THE BDM CORPORATION

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COAL GASIFICATION SYSTEMS ENGINEERING AND ANALYSIS

FINAL REPORT

VOLUME II

December 31, 1980

BDM/H-80-800-TR

This Technical Report is submitted to George C. Marshall Space Flight Center under Contract Number NAS8-33824.

"LEGAL NOTICE"

FOREWORD

This final report is submitted to the George C. Marshall Space Flight Center (MSFC), National Aeronautics and Space Administration, by The BDM Corporation, Suite 32, Holiday Office Center, 3322 Memorial Parkway SW, Huntsville, Alabama, 35801, as fulfillment of the final report requirement of Contract Number NAS8-33824, entitled "Coal Gasification System Engineering and Analysis."

Mr. Thomas Irby is the MSFC Contract Officer Representative. This study is to provide MSFC a basis for their support of the Tennessee Valley Authority Coal Gasification Project, consisting of a four 5,000 ton/day module coal gasification facility. Major project support for this study is provided by the Mittelhauser Corporation acting as a subcontractor.

Dr. Jerry V. Fox is The BDM Corporation program manager. Mr. M. Dale Dowden is the Mittelhauser Corporation project manager.

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BDM Corporation technical staff are Dr. R. M. Bass, Mr. C. Carter, Mr. M. F. Funke, Mr. S. Majied, Dr. B. S. Morgan, Mr. J. R. Query and Dr. J. M. Siegel. Mittelhauser Corporation technical staff are Mr. T. A. Atkins, Mr. T. W. Barrs, Mr. W. C. Chambers, Mr. S. E. Heffley, and Mr. S. H. McFeely. Consultants were Mr. M. R. Beychok and Mr. Henry Ho.

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CHAPTER I
INTRODUCTION

A. BACKGROUND

The purpose of this study was to support the feasibility analysis and systems engineering studies for a 20,000 tons per day medium BTU (MBG) coal gasification plant to be built by TVA in Northern Alabama. TVA plans to build the plant in four modules of 5000 tons per day each with the first module on line in mid 1985. In this study, The BDM Corporation and its subcontractor, the Mittelhauser Corporation, have provided assistance to NASA Marshall Space Flight Center for its feasibility analyses and systems engineering studies in support of the TVA project.

As part of its feasibility analysis, TVA has contracted with three engineering firms for preliminary plant designs based on five different gasifiers. These designs will be used to select a gasifier or gasifiers for the plant.

B. OBJECTIVES, ASSUMPTIONS, GUIDELINES AND LIMITING FACTORS

The major objectives of the study were as follows:

1) Provide design and cost data to support the selection of a gasifier technology and other major plant design parameters.
2) Provide design and cost data to support alternate product evaluation (methane, methanol, gasoline, hydrogen).
3) Prepare a technology development plan to address areas of high technical risk.
4) Develop schedules, PERT charts, and a work breakdown structure to aid in preliminary project planning.

Assumptions, guidelines and limiting factors are summarized briefly in Figure I-B-1. Detailed guidelines were provided in a TVA publication, "Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," March 1980.

Other Items specified in the TVA document include the following:

Location: Murphy Hill, Alabama

Coal: Kentucky No. 9

Coal cost: $1.25/mm Btu; 1/1/80 dollars

Product Gas:

Pressure:	600 psig minimum
Temperature:	120 degrees F maximum
Higher Heating Value:	285 Btu/SCF minimum
Total Sulfur:	200 ppm maximum
Total Moisture:	7 lbm/MMSCF maximum
Chemical Composition:	Within the constraints described above, the composition of the gas at the plant fence may be established solely by the coal gasification and gas cleanup processes.

Design Capacity: 20,000 tons of coal per day, in four modules of 5000 tons per day each

On stream Factor: 90 percent

Module life: 20 years after startup

Initial Operation Schedule:

First module	6/1/85
Second module	6/1/86
Third module	1/1/87
Fourth module	6/1/87

Candidate Gasifiers:
- Koppers-Totzek
- Texaco
- Babcock and Wilcox
- Lurgi
- BGC/Lurgi

Figure I.B.1. Major Gasification Plant Parameters

- Site and Transportation Conditions
- Coal Receiving and Handling
- Building and Support Structures
- Codes and Standards
- Coal and Water Characterization
- Byproduct Specifications and Disposition
- Environmental Control Guidelines
- Detailed Economic Assumptions
- Cost Power; Construction and Escalation rates for Operations and Maintenance Labor

C. STUDY APPROACH AND MAJOR RESULTS

The investigative flow and major study results are illustrated in Figure I-C-1. As a baseline for all tasks, the major design-related features of each generic plant system were characterized in a "catalog." A facility requirements document providing plant specifications for design guidance was developed jointly with NASA. Based on the catalog and requirements data, approximately 17 designs and cost estimates were developed for MBG and alternate products. Additionally, a series of generic trade studies was conducted to support all of the design studies.

To supplement the designs, a set of cost and programmatic analyses were conducted. The cost methodology employed for the design and sensitivity studies was documented and implemented in a computer program. Plant design and construction schedules were developed for the K-T, Texaco and B&W MBG plant designs. A generic Work Breakdown Structure was prepared, based on the K-T design, to coincide with TVA's planned management approach. An extensive set of cost sensitivity analyses was completed for the K-T, Texaco and B&W design. Product price competitiveness was evaluated for MBG and the alternate products. Finally, a draft Management Policy and Procedures Manual developed by TVA was evaluated and modifications were recommended.

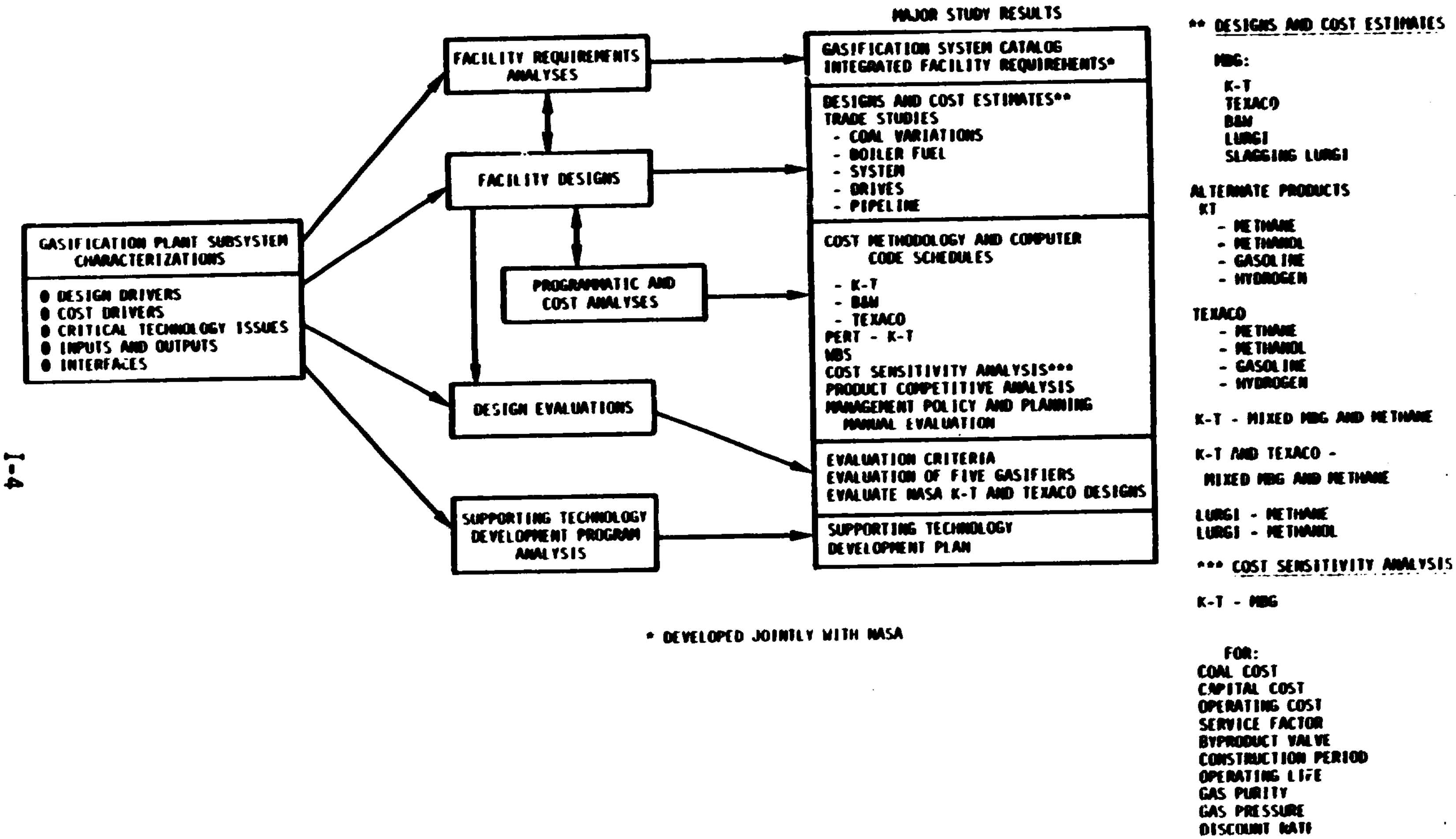


Figure I.C.1. Program Organization

Several evaluation tasks were conducted. Evaluation criteria were developed for assessing the preliminary gasifier designs prepared for TVA by three engineering firms. An evaluation of the advantages and disadvantages of the five candidate gasifiers was prepared. Finally, NASA's own K-T and Texaco designs were compared to the BDM/Mittelhauser designs.

A supporting technology development plan was developed to address high technology risk issues. The issues were identified and ranked in terms of importance and tractability, and a plan developed for obtaining data or developing technology required to mitigate the risk.

D. ORGANIZATION OF THIS REPORT

Each of the major study results listed in Section C is described in the remainder of this report. The following outlines the report by chapters.

Chapter II	Gasification System Characterizations
Chapter III	MBG Facility Designs
Chapter IV	Trade Studies
Chapter V	Cost Analyses and Methodology
Chapter VI	Alternate Product Designs
Chapter VII	Schedule and Network Analysis
Chapter VIII	Product Competitive Evaluations
Chapter IX	Work Breakdown Structure
Chapter X	Management Policies and Procedures
Chapter XI	Commercial Design Assessment
Chapter XII	Assessment of Critical Technology Needs

In addition, complete results of each of the project tasks are included as Appendices A through H.

(1) Appendix A Coal Gasification System Catalog
(2) Appendix B Medium Btu Gas Design
(3) Appendix C Alternate Product Designs
(4) Appendix D Costs and Economic Studies
(5) Appendix E Methodology of Cost Determination
(6) Appendix F Critical Technology Evaluation and Recommendations
(7) Appendix G Commercial Design and Technology Evaluation
(8) Appendix H Work Breakdown Structure

CHAPTER II
GASIFICATION SYSTEM CHARACTERIZATIONS

This chapter briefly describes the gasification technologies, major design and cost considerations, and the other systems in the gasification plant. Raw materials required for plant operation and byproduct markets are also discussed briefly. These topics are treated in more detail in Appendix A.

A. DESCRIPTION OF GASIFICATION TECHNOLOGIES

1. Introduction

TVA selected five gasification technologies for evaluation: Koppers-Totzek, Texaco, Lurgi Dry Ash, Slagging Lurgi, and Babcock and Wilcox. Each of these is described below. The Unit Operations referenced in these descriptions are discussed in Section D below and in Appendix A. Unit operation numbers referenced are from the system/unit operation correlation used in the catalog in Appendix A.

2. Koppers-Totzek

The Koppers-Totzek gasifier is a high temperature, cocurrent entrained flow gasifier which accepts coal from Coal Preparation along with oxygen and steam to produce intermediate BTU gas. It is a proprietary unit licensed by Krupp-Koppers of Germany. Sized coal 1/4"x0 from Coal Preparation, Unit Operation 11, enters the pretreatment area of Gasification, Unit Operation 20, where it is crushed and ground to 70% minus 200 mesh, and dried to 2% moisture. It is then fed to eight screw conveyors that feed four pairs of burners located 90° apart. There are four feed points on the four--headed gasifier with 2 burner heads at each point. Each burner projects a jet to converge with the line of discharge of the other. Oxygen from Air Separation, Unit Operation 80, and steam from Steam Generation, Unit Operation 84, carry the coal through the burners into the gasifier.

The oxygen, steam, and coal react to gasify the carbon and volatile matter of the coal and to convert the coal ash into molten slag

which is sent to Solids Treatment, Unit Operation 31. The gas exiting each gasifier is direct water quenched to below the ash fusion temperature, in order to solidify entrained slag droplets. The remaining slag forms a layer on the refractory walls and flows down through a separate chute into quench tanks.

Low pressure steam is produced in the water jackets of the gasifiers from waste heat that passes through the refractories.

After the gas is quenched, gas and entrained ash particles pass through a waste heat boiler where the gas is cooled to approximately 350^{0} by raising high pressure steam. The gas is then scrubbed for particulate removal. The clean intermediate BTU product gas is then further cooled in Gas Cooling, Unit Operation 21, before going to Acid Gas Removal, Unit Operation 22.

With the K-T gasifier, as with all high temperature entrained flow gasifier, no tars, phenols, oils, etc., are produced so the gas requires less cleanup that those systems that produce hydrocarbons. Because of the high operating temperatures the gasifer requires an appreciable amount of oxygen per pound of coal fed. The K-T gasifier requires about 0.9-1.0 lb O_2 per lb coal fed. Steam consumption is approximately 0.4 lb per lb of coal. The higher heating value of the dry gas produced from the K-T gasifier is in the range of 285-300 BTU/SCF. The Koppers-Totzek gasifier typically operates at a pressure of about 7 psig. Maximum temperatures can run as high as 3300^{0}F.

3. Texaco

The Texaco Coal Gasification Process uses a coal slurry feed, consisting of fresh ground coal together with recycled fine slag and carbon with a total solids content 50 to 65% by weight. The slurry is pumped from mix tanks in the grinding and slurry section to the gasifier slurry tank. A circulating pump circulates the slurry through this tank and supplies slurry to the suction of the high pressure charge pump.

The coal-water slurry is fed through a specially developed burner into a refractory-lined gasifier reactor. Partial combustion with oxygen takes place at a pressure of 600 psig, or higher, and a

temperature in the range of 2300 to 2800°F to produce a gas consisting mainly of CO, H_2, CO_2, and steam. Most of the sulfur in the coal is converted to H_2S, and the balance converts to COS. Nitrogen and argon from the oxygen feed appear in the gas together with most of the nitrogen from the coal. The gas contains a small amount of methane, some unconverted carbon and all of the ash in the form of slag. The gas is essentially free of uncombined oxygen.

The upper section of the gasifier is the refractory-lined chamber in which the partial oxidation reaction takes place. In many conceptual designs, part of the gas is withdrawn and cooled to below the ash fusion point by mixing with cooled recycle gas. Entrained slag particles, solidified by cooling, are then removed from the gas. The gas is then cooled by raising high-pressure steam in a specially-designed waste heat boiler. The gas then passes to the Gas Cooling System, Unit Operation 21. To date, these high-pressure steam generators have not been commercially proven in coal gasification service.

At least a portion of the gas from the gas generator reaction section passes straight down into the quench section of the gasifier. This stream carries the bulk of the larger particles of slag, and it is immediately quenched with water from the 2300 to 2800°F range to about 400°F. The gas from the generator quench chamber joins the main stream of gas going to the gas cooling operation.

Water from the gasifier quench chamber is cooled and combined with water from the carbon-scrubber lower section. Both streams contain fine slag and unconverted coal. The water steam from the lock hopper and water from the final product cooler separator join this steam, and the total flows into the flash pot.

In the clarifier, the fine slag and unconverted coal settle out, leaving a clarified water overflow that is pumped back to the carbon scrubber via the gray water drum. Makeup water is added at this point.

The clarifier underflow is fed to a centrifuge for dewatering. The concentrated underflow is returned to the coal grinding and slurry-

ing section and mixed with coal-feed slurry. The filtrate is returned to the clarifier.

Most of the ash in the coal feed agglomerates into essentially carbon-free molten slag droplets, which are quenched and solidified in the lower quench section of the reactor. This slag is settled through the quench water into the lock hopper. The lock hopper is periodically dumped onto a screen, from which the slag is conveyed to the solids treatment system.

4. Lurgi

The Lurgi gasifier, dry ash, gravitating bed type, is commercially available from Lurgi Kohle and Mineraloeltechnik. The gasifier is a water jacketed pressurized unit comprised of a series of vertically stacked vessels. There are, from top to bottom, a coal hopper, coal lock, water jacketed gasifier, ash lock, and ash quench chamber.

Coal is conveyed from Coal Preparation, Unit Operation 11, to the coal hopper from which it is fed by gravity to the depressurized coal lock through a hydraulically operated valve. The lock is then isolated and pressurized with a slipstream of inert gas (mainly N_2) and the coal is tranferred to the gasifier through another hydraulically operated valve. The empty lock is isolated, depressurized through a bag filter and vented either to the atmosphere or the Incinerator, Unit Operation 41. The gas displaced from the coal and lock hoppers during loading is routed similarly. Coal dust recovered in the filter is returned to the coal hopper.

The coal flowing down through the gas produced represents a slowly moving bed which has several distinct zones. In the first zone at the top of the gasifier, coal is preheated and dried by contact with the hot crude gas leaving the reactor. As the coal moves down and is heated further, devolatilization occurs and gasification commences. The bottom of the bed is a combustion zone where carbon reacts with oxygen to form CO and CO_2. The oxidation provides the overall heat for the gasification and devolatilization reactions which are endothermic. Only a negligible amount of unburned carbon remains in the ash.

When MBG is to be made, oxygen from Air Separation and Oxidant Feeding, Unit Operation 80, and steam enter the gasifier near the bottom and are heated as they rise upward to the combustion zone by the hot ash moving down from the combustion zone. Oxygen flow rate is controlled to accomplish complete gasification of coal. Steam rate is controlled to maintain a specified gasifier bottom temperature to prevent melting or clinkering of the ash.

A portion of the gasifier process steam is generated at about the operating process of the gasifier, in the gasifier jacket. The balance is provided through waste heat recovery or from Steam Generation, Unit Operation 84.

The crude gas leaving the gasifier contains appreciable quantities of tars, oils, naptha, phenols, fatty acids, ammonia, hydrogen sulfide, sulfur compounds, and a small amount of coal and ash dust. The crude gasifier effluent temperature ranges from 575^{o}F to over 1000^{o}F. The effluent flows through a scrubbing cooler where it is washed with a stream of process condensate. The washing process quenches the gas to about 350-400^{o}F and condenses the high boiling tar fractions. Coal and ash dust are removed with the condensed tar leaving the quenched effluent gas essentially free of particulate matters.

Ash from the process is continuously collected by a rotating ash grate and moved to the ash lock hopper. Ash collected in the lock is depressurized and discharged batchwise to an ash quench chamber where it is cooled in water. The ash lock is pressurized with steam.

The abrasive slurry from each gasifier train flows to a common transfer tank using water as the motive fluid. Ash grinders are provided to prevent large chunks of slag from plugging transfer lines. The ash slurry is then sent to Solids Treatment, Unit Operation 31.

5. <u>Babcock and Wilcox</u>

The Babcock and Wilcox gasifier is a high temperature, cocurrent entrained flow gasifier which accepts coal from Coal Preparation along with oxygen and steam to produce medium BTU gas. It is a proprietary unit licensed by Babcock and Wilcox.

Sized coal 1/4"x0 from Coal Preparation, Unit Operation 11, enters the pretreatment area of Gasification, Unit Operation 20, where it is pulverized to 70% minus 200 mesh and tangentially injected through two rows of water cooled nozzles into the gasifier. Recycled char from the gasifier outlet gas cyclones is also injected through water cooled nozzles in to the bottom row of burners. Both the coal and char are fired with oxygen from Air Separation, Unit Operation 80. The coal and char are partially combusted to form a hot reducing gas. At the high temperatures present in the gasifier, the ash in the coal and tar becomes molten and continuously flows down the walls of the gasifier to the slag tap hole. From the tap hole, the slag enters a water quench tank where it is cooled. From the quench tank it flows to the Solids Treatment System, Unit Operation 31.

In the gasification section, there is an inner shell of water cooled tubes (water wall) where saturated steam is produced. In the hot reaction zone, the tubes are covered with a dense refractory suitable for contact with molten flowing slag. Above the reaction zone, the tubes are bare for greater radiation cooling prior to entrance into the waste heat boiler section.

The gas exits the gasifier proper at about 1800^{o}F and enters the waste heat boiler section where it is cooled to 700^{o}F. From the waste heat boilers, the gas enters a cyclone where 90-95% of the carry-over ash and char is removed. This char and ash stream, as mentioned previously, is injected back into the gasifier. The 700^{o}F gas is further cooled and cleaned in Gas Cooling, Unit Operation 21, before going to Acid Gas Removal, Unit Operation 22.

6. <u>Slagging Lurgi</u>

This system is divided into the following four subsystems:

- Coal and Flux Feed
- Gasification
- Raw Gas Treating
- Slag Handling.

a. Coal and Flux Feed

Flux is an agent which forms a eutectic mixture with the coal ash in the gasifier, lowering its melting point to make slag formation easier. It is shipped to the facility from outside sources.

The coal and flux are mixed in Coal Preparation, Unit Operation 11, and then fed to the coal bunkers by a belt conveyer system. The feed chutes at the bottom of the coal bunkers control the flow of coal into the coal locks. Each gasifier has two coal locks that operate automatically on a cyclic basis. There, coal locks are pressurized with mostly N_2 and alternately feed the coal surge vessel.

b. Gasification

The design of the gasifier is based on proprietary technology held by Lurgi Kohle Mineraloeltechnik and the British Gas Corporation. It is similar to the dry-ash Lurgi gasifier described earlier, except that in the bottom of the gasifier the coal ash melts as a eutectic with the added flux to form slag. The molten slag collects at the bottom and is removed intermittently from the gasifier through a slag tap hole.

The coal and flux, entering the top of the Gasifier, descends in a moving bed in countercurrent flow to steam, oxygen and produced gas. While traveling from the top to the bottom of the gasifier, the coal is dried, devolatilized, and gasified. The heat required for these three steps is supplied by the exothermic reaction between the carbon in the coal and the oxygen in the bottom of the gasifier.

As the produced gas passes through the coal bed, its final composition is determined by the following:

- Exothermic and endothermic reactions occurring simultaneously in the gasification zone.
- Formation of hydrocarbons, phenols, fatty acids, and minor organic compounds in the devolatilization zone.
- Evaporation of coal moisture in the drying zone.

c. Raw Gas Treating

Raw gas is treated similarly to that from a dry-ash Lurgi gasifier, as described earlier.

d. Slag System

After the coal ash melts as a eutectic with the added flux to form slag, the molten slag collects at the bottom of the gasifier and is tapped intermittently through a tap hole into the Quench Vessel. In the Quench Vessel, the slag granulates immediately upon contact with the quench water. The granulated slag falls into the Slag Hopper and is dumped once or twice an hour. The slag and water mixture that is dumped goes to Solids Treatment, Unit Operation 31.

B. DESIGN AND COST DRIVERS

Major design and cost drivers, developed for each major plant system, are presented in detail in Appendix A. Design drivers are the specifications or other considerations that are major determinants of the resulting design. Cost drivers are the major determinants of product cost.

For the plant as a whole, the major design drivers are plant capacity; coal characteristics (carbon, hydrogen, sulfur, trace elements, moisture); product specifications (type of products, pressure, sulfur level); and waste water effluent restrictions. Plant capacity establishes the scale for the design, and will have a major impact on solids handling, utility scaling, train configurations, and sparing. The coal characteristics will affect the choice of gasifier and will drive design of all cleanup systems. Product specification will determine requirements for compression and sulfur removal. If the product is not MBG, product specifications may affect the choice of gasifier and will determine downstream processing requirements. Water effluent specifications will have a major impact on design of the complex waste treatment systems.

The major cost drivers are capacity, coal characteristics, product specifications, and coal cost. The capacity will determine the applicable scale economies. Coal characteristics and product specifications will determine the product yield and selection of major capital items

(gasifier, gas cleanup, compression, conversion). The coal cost is a major operating cost independent variable, while labor and spare parts are determined primarily by capital costs.

C. GASIFICATION FACILITY SYSTEMS

1. Introduction

The coal gasification facility comprises about 25 major systems or types of unit operations, listed in Figure II.C.1. To provide an overview of the facility, the nature and purpose of each system are described briefly in this section. A detailed description of each process is provided in Appendix A. Figure II.C.2. characterizes the major system components regarding cost, lead time, and technical uncertainty.

2. Coal Receiving, Storage, and Transfer

Unit Operation Number 10

The Coal Receiving, Storage, and Transfer System provides for the unloading of coal delivered to the plant either by barge or truck, transporting the coal to storage, reclaiming the coal from storage, reducing the size of the coal, and transporting the coal to Coal Preparation, Unit Operation 11.

3. Coal Preparation

Unit Operation Number 11

This system receives raw coal from Coal Receiving, Storage, and Transfer, Unit Operation 10; reduces the coal to the proper size; screens out and recycles the oversize fraction; and transfers the properly sized coal to Gasification, Unit Operation 20, for direct gasification or further treatment.

In some gasifiers, such as Lurgi, the Coal Preparation System is not part of the gasifier licensor's proprietary technology. In others, such as Texaco, final crushing and slurrying are part of the proprietary package, while in the Koppers-Totzek process all coal crushing, drying and feeding are considered proprietary by the licensor.

NUMBER	
10	COAL RECEIVING, STORAGE AND TRANSFER
11	COAL PREPARATION AND FEEDING
20	GASIFICATION
21	GAS COOLING
22	ACID GAS REMOVAL
23	COMPRESSION
31	SOLIDS TREATMENT SYSTEM
32	TAR-OIL SEPARATION
33	PROCESS CONDENSATE TREATMENT
34	PHENOL RECOVERY
35	AMMONIA RECOVERY
36	SULFUR RECOVERY
37	BIOLOGICAL TREATMENT
39	COOLING WATER SYSTEM
41	INCINERATION
80	AIR SEPARATION AND OXIDANT FEEDING
81	FINAL SOLIDS DISPOSAL
82	BY-PRODUCT STORAGE AND LOADING
83	SULFUR STORAGE AND LOADING
84	STEAM GENERATION
85	RAW WATER TREATMENT
86	FLUE GAS TREATMENT
87	PLANT ELECTRICAL SYSTEM
88	BUILDINGS AND SUPPORT FACILITIES
89	CONTROL AND INSTRUMENTATION

Figure II.C.1. Unit Operation Categories

	FIELD FABRICATED VESSELS	SHOP FABRICATED VESSELS	COMPRESSORS BLOWERS	EXCHANGERS	PUMPS	FIRED EQUIPMENT	SOLID TRANSFER	SOLID STORAGE	FILTRATION	CLARIFICATION	SPECIALTY EQUIPMENT	SIZE REDUCTION CLASSIFICATION	TANKAGE
10 COAL RECEIVING STORAGE & TRANSFER	0	4	4	0	4	0	2	4	0	0	1 2	0	0
11 COAL PREPARATION & FEEDING	4	1 3 4 1 1	4	0	0	4	2	4	0	0	4 [illegible]	0	0
20 GASIFICATION (NOTE 1)	0	3 3 3 3 3	0 2 4 0 0	0 3 2 0 0	4	4 0 0 0 0	3 4 4 4 4	4 0 0 4 4	0 4 0 0 0	0 4 0 0 0	0 3 0 2 2	0	0
21 GAS COOLING (NOTE 1)	0	4	0	4 3 4 3 3	4	0	0	0	0	0	0	0	0
22 ACID GAS REMOVAL (NOTE 2)	1/4/4/4	1/2/1/1	1/1/0/4	1/4/1/1	1/4/4/4	0	0	4	4/4/4/4	0	0/1/4/1	0	4/4/4/4
23 COMPRESSION	0	0	2	0	0	0	0	0	0	0	0	0	0
31 SOLIDS TREATMENT SYS	0	4	0	4	3 2	0	3	3	3 2	0	2	0	4
32 TAR OIL SEPARATION	0	4	0	0	4	0	0	0	0	0	4	0	0
33 PROCESS CONDENSATE TREATMENT	0	1 2 3	0	2 3	2	0	4	0	0	0	4	4	4
34 PHENOL RECOVERY	4	1 2 3	0	4	4	0	0	0	0	0	1 2 3	0	4
35 AMMONIA RECOVERY	0	1 2 3	0	4	4	0	0	0	0	4	0	0	4
36 SULFUR RECOVERY	0	1 2	4	4	2	0	4	0	0	4	1 2	4	4
37 BIOLOGICAL TREATMT	0	0	0	0	4	0	4	4	4	4	4	0	0
39 COOLING WATER SYSTEM	4	0	0	4	4	0	0	0	4	0	3	0	0
41 INCINERATION	0	0	4	4	0	4	0	0	0	0	4	0	0
44 AIR SEPARATION AND OXIDANT FEEDING	0	2	1	4	4	0	0	0	0	0	0	0	1 2
81 FINAL SOLIDS DISPOSAL	0	0	0	0	3	0	1	3	0	0	4	0	4
82 BY PRODUCT STORAGE AND LOADING	0	0	4	4	4	0	0	0	0	0	4	0	4
83 SULFUR STORAGE AND LOADING	0	0	0	0	1	0	0	0	0	0	1	0	4
84 STEAM GENERATION	0	0	4	4	4	1	4	4	0	0	0	1	0
85 RAW WATER TREATMENT	4	1	0	1	1	0	4	4	4	4	1 2	0	1
86 FLUE GAS TREATMENT	1 2 3	1	1 2	3	2 3	2 3	4	4	4	4	1 2 3	0	4
87 PLANT ELECTRICAL SYS	0	0	0	0	0	0	0	0	4	0	1 2	4	0
88 BUILDING & GENERAL SUPPORT FACILITIES	0	4	4	4	4	0	1	0	0	0	1 2	0	0
89 CONTROL & INSTRUMENTATION (NOTE 3)													

CHARACTERIZATION CODES
0 NOT INCLUDED IN SYSTEM
1 HIGH COST ITEM
2 LONG LEAD TIME ITEM
3 CRITICAL TECHNOLOGY AREA
4 NON CRITICAL ITEM

NOTES
(1) NUMERICAL SEQUENCE REPRESENTS THE FOLLOWING ORDER OF GASIFIERS: B&W, TEXACO, K-T, LURGI, BGC LURGI
(2) NUMERICAL SEQUENCE REPRESENTS THE FOLLOWING ORDER OF AGR UNITS: SELEXOL, STRETFORD, BENFIELD, RECTISOL
(3) NOT IDENTIFIED IN SOURCE REFERENCE LITERATURE (NO DATA REPORTED)

Figure II.C.2. System Component Characteristics

4. Gasification

Unit Operation Number 20

The gasification section produces a hot raw gas which is fed to raw gas cooling, Unit Operation 21. The gasifier is fed with sized coal or coal slurry from Unit Operation 11. Other feeds are oxygen from Unit Operation 80 and in some cases steam from Unit Operation 84. The gasifier may produce ash and slag which are sent to Solids Treatment, Unit Operation 31.

The five gasification technologies evaluated in this study are Koppers-Totzek, Texaco, Lurgi Dry Ash, Slagging Lurgi, and Babcock and Wilcox. These are described in Section A above.

5. Raw Gas Cooling

Unit Operation Number 21

The purpose of this unit is to cool the gasifier effluent gas to more amenable processing temperatures in subsequent systems and to separate ash and solids from the gas. This system description is dependent upon gasifier selection, as some gasifiers include one or more stages of cooling as an integral part of the gasifier system.

The portions of gas cooling associated with the gasifier are proprietary and are handled by the gasifier licensor. The remaining portions of gas cooling are non-proprietary.

6. Acid Gas Removal

Unit Operation Number 22

The purpose of the Acid Gas Removal System is to remove H_2S, other sulfur compounds, and CO_2 from Raw Gas Cooling, Unit Operation 21. Potentially applicable commercial processes include Selexol, Stretford, Benfield, and Rectisol.

The Stretford Process is a proprietary process licensed by North West Gas Board, Ltd., and offered by Parsons and others in the U.S.

Selexol is a proprietary process developed and licensed by Allied Chemical Corporation.

Rectisol is a proprietary technology licensed by Lurgi Kohle and Mineraloeltechnik, GmbH.

Benfield is a proprietary process developed and licensed by Benfield Corporation, a subsidiary of Union Carbide Corporation.

The processes vary in attainable gas purity, operating pressure and cost.

7. Compression

Unit Operation Number 23

The purpose of this unit is to compress and dry the gas produced for delivery to the pipeline at 600 psig. The dry compressed gas is usually metered before delivery to the pipeline.

The compressors may be located at various positions within the overall processing sequence. Regardless of where the gas is compressed, the drying unit must be the last unit in the gas processing sequence.

8. Solids Treatment System

Unit Operation Number 31

The purpose of this unit is to collect and dewater the various solids slurries, or sludges resultant from the Facility Operation for economical, environmentally acceptable disposal.

The processes involved in this system are non-proprietary and are supplied by various US vendors.

Typically, ash and slag from the gasifier and gas cooling system, biological sludges, and solid wastes from process condensate treatment are treated in this unit.

9. Tar-Oil Separation

Unit Operation Number 32

This unit is used in the Lurgi and BGC-Slagging Lurgi Processes to collect the condensate from Gasification, Shift Conversion, Gas Cooling, and Rectisol. The condensate, called gas liquor, contains tar, oil, dust, and other impurities. Gravity settling tanks are used to separate the tar, oil, and dust from the gas liquor. The tar and dust are recycled to Gasification so there is no net production. The oil is recovered either as a saleable product or for plant fuel.

This design is based on proprietary technology held by Lurgi Kohle and Mineraloeltechnik, GmbH.

10. Process Condensate Treatment

Unit Operation Number 33

The purpose of this unit is to collect and treat all facility liquid effluent streams. The facility design is predicated on "zero discharge." A wastewater treatment system permitting recycle and reuse of treated water would be required to meet this requirement.

This system contains various non-proprietary processes that can be supplied by several US vendors.

The number, type, quantity, and composition of liquid effluents generated within the facility are dependent on the gasifier technology selected and the process systems selected to produce the end product(s). The following list identifies the possible liquid effluents to be treated and, in some cases, may be mutually exclusive:

- Oily Water Sewers
- Coal Pile Run Off
- Storm Water Run Off
- Demineralizer Regenerant Wastes and Rinse Water
- Cooling Tower Blowdown
- Boiler (Steam Generator) Blowdown
- Ammonia Recovery System Blowdown
- Rectisol Blowdown
- Sanitary Waste Water
- Flue Gas Treatment Slurry
- Gasifier Slag Quench Drains
- Separated Water From Solids Treatment
- Filtrate From Biological Treatment.

11. Phenol Recovery - Phenosolvan

Unit Operation Number 34

The Phenosolvan process is used to treat phenol-containing water from Process Condensate Treatment, Unit Operation 33. Phenol is

recovered for sales, and dephenolized water is transferred to Ammonia Recovery, Unit Operation 35, for further processing.

The Phenosolvan process is a proprietary process licensed by and offered by American Lurgi Corporation (N.Y.).

12. Ammonia Recovery

Unit Operation Number 35

This unit receives sour water from Phenol Recovery, Unit Operation 34 (if present), containing volatile components such as ammonia, carbon dioxide, hydrogen cyanide, and hydrogen sulfide. The feed enters a stripper where the volatile components are stripped.

Anhydrous ammonia of commercial purity (less than 10 ppm hydrogen sulfide) is produced and sent to By-Products Storage, Unit Operation 82, for sales.

There are two processes of interest, the Phosam-W process is licensed by USS Engineers, Inc. The CLC process is a proprietary process licensed by the American Lurgi Corporation.

13. Sulfur Recovery - Claus Sulfur

Unit Operation Number 36

The purpose of this unit is to recover elemental sulfur from acid gas from Acid Gas Removal, Unit Operation 22. Acid gas is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H_2S acid gases. The chemistry of the process involves converting the H_2S to elemental sulfur. A tailgas waste stream is also produced.

Several commercial processes are available for reducing the sulfur content of sulfur recovery unit tail gas to an environmentally acceptable level. The Beavon sulfur removal process is capable of reducing the sulfur content in the tail gas to less than 100 ppm. It is a proprietary process licensed by Ralph M. Parsons Co.

14. Biological Treatment - POAS Process

Unit Operation Number 37

The purpose of this unit is to treat effluents from sanitary waste treatment and the process condensate treating system, Unit Operation

33, to such an extent that they can be recycled into the facility water system.

This is a proprietary process licensed by Union Carbide Corporation.

15. Biological Treatment - Air Activated Sludge Process

Unit Operation Number 37

The purpose of this unit is to remove organic contaminants from plant wastewater. In addition to organics, biological oxidation removes some amounts of trace metals, and trace organics. It also removes phenols, cyanides, and ammonia that are present in coal conversion wastes.

16. Cooling Water System

Unit Operation Number 39

The purpose of this unit is to provide cooling water to the various process users in the facility.

This is a non-proprietary system and is supplied by several US vendors.

17. Incineration

Unit Operation Number 41

This unit combusts environmentally objectionable constituents in various vent gases and waste gases from facility systems and renders these gases into a form that is acceptable for release into the atmosphere.

18. Air Separation and Oxidant Feeding

Unit Operation Number 80

The purpose of this unit is to supply oxygen to the gasifiers. The gaseous oxygen stream is usually between 95 and 98 percent pure by volume, and it is produced by distillation of liquified air.

This is a proprietary but non-licensed process offered by several designer/manufacturers in the US.

19. Final Solids Disposal

Unit Operation Number 81

The purpose of this unit is to store solid waste generated by facility operation during the facility life.

This is a non-proprietary system.

20. <u>By-Product Storage and Loading</u>

Unit Operation Number 82

The purpose of this system is to receive and store various by-products from recovery units, transfer the by-products to a loading area and into the proper vehicles for transportation.

This system will receive and store: (1) ash residue from Coal Preparation and Feeding, Unit Operation 11; (2) oils and tars from Tar-Oil Separation, Unit Operation 32; (3) liquid anhydrous ammonia from Ammonia Recovery, Unit Operation 35; and (4) phenol from Phenol Recovery, Unit Operation 34.

21. <u>Sulfur By-Product Storage</u>

Unit Operation Number 83

The purpose of this unit is to store elemental sulfur of high purity from Sulfur Recovery and Tail Gas, Unit Operation 36. The sulfur flows by gravity to sulfur pits. From the pits it is pumped as a 280°F liquid to storage tanks awaiting loading on a barge or truck. This is a non-proprietary system available from a variety of U.S. vendors. Alternately, sulfur may be solidified by a technique such as prilling or flaking.

22. <u>Steam Generation</u>

Unit Operation Number 84

The purpose of the steam generation unit is to provide high pressure super-heated steam to supplement the steam generated by waste heat recovery in the Gasifier Unit Operational Cooling Systems, Unit Operation 21.

The steam generators are water tube boilers fired either by raw coal, coal fines or MBG.

This is a non-proprietary process, and the equipment is supplied by several U.S. vendors.

23. <u>Raw Water Treatment</u>

The Raw Water Treatment Unit is typically designed to provide treated and untreated water for the following facility water systems:

- Fire Water
- Service Water

- Potable Water
- Cooling Water
- Boiler Feedwater

The processes by which the raw water is treated for the above services are non-proprietary and are supplied by several US vendors.

Raw water is usually pumped from the river to a Fire Water-Raw Water Storage tank (or pond).

The Raw Water-Fire Water Storage Tank provides surge capacity for Water Treatment as well as storage capacity for fire water. During an emergency, fire water is pumped from the tank to the fire water header system.

24. Flue Gas Treatment

Unit Operation Number 86

The purpose of this system is to clean up effluent gas containing sulfur dioxide and particulates from Steam Generation, Unit Operation 84. The Wellman-Lord sulfur dioxide recovery process is a proprietary process licensed by Davy Powergas, Inc., Houston, TX.

The Double Alkali process is offered by several US vendors including Combustion Engineering Associates and FMC Corporation.

25. Plant Electrical System

Unit Operation Number 87

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

- Develop the necessary voltage stepdown arrangement for plant requirements
- Distribute the necessary power to the plant equipment.

This is non-proprietary equipment and is supplied by several US vendors.

26. Buildings and Support Facilities

Unit Operation Number 88

The purpose of this unit is to provide equipment or services to support the Gasification Facility at the facility level.

The equipment and services provided in this unit are non-proprietary.

This unit is a general facility category and would typically provide the following equipment or service:

- Administration Building
- Laboratories
- Change Rooms
- Warehouses
- Maintenance Buildings
- Operation Centers
- Security Offices
- First Aid Facility
- Fire House
- Visitor Reception
- Plant Fencing
- Plant Lighting
- Roads, Bridges, and Sewers
- Docking Facilities
- Interconnection Pipe Ways
- Fire Protection Network
- Flare Stacks and Headers
- Plant Instrument Air Compressors
- Environmental Monitoring
- Site Preparation.

27. Control And Instrumentation

Unit Operation Number 89

The purpose of this unit is to provide operational control of the facility and supervisory master control of the facility module operation.

This item has not been identified as a discrete cost or operational center in the studies or evaluation presented in the literature.

A separate subcontract is often executed with a supplier of computer data acquisition systems, to provide a data acquisitions system for a facility and to interface it with the instrumentation and control loops within the facility.

D. SYSTEM CHARACTERIZATION

The systems employed, the nature of their interconnections, and stream characteristics depend on the gasifier technology and the specific plant design. A representative example of a system configuration for the Lurgi gasifier with major streams identified, is shown in Figure II.D.1. This configuration contains all of the systems described in Section C above. A detailed description of all typical stream components, pressure, and temperatures ranges are provided in Appendix A. Detailed flow sheets and stream characteristics are provide in Chapter IV and Appendix B.

E. RAW MATERIALS ANALYSIS

Most of the unit operations identified in this study require raw materials to support the process function. These raw materials are necessary for unit performance and represent a cost of operation both annually and as an initial cost.

This section presents the results, summarized in Figure II.E.1., of this analysis. The approach taken is to present all raw material elements that are required by all the gasification facility systems. Therefore, dependent upon the final design configuration, only some of the raw material listed will be required to support the selected unit operations for the TVA Coal Gasification Facility.

The data are presented in a matrix format. They include a list of raw materials, the system(s) with which they are associated, both the initial and replacement quantities required, the unit cost, a commercial source, and transportation information.

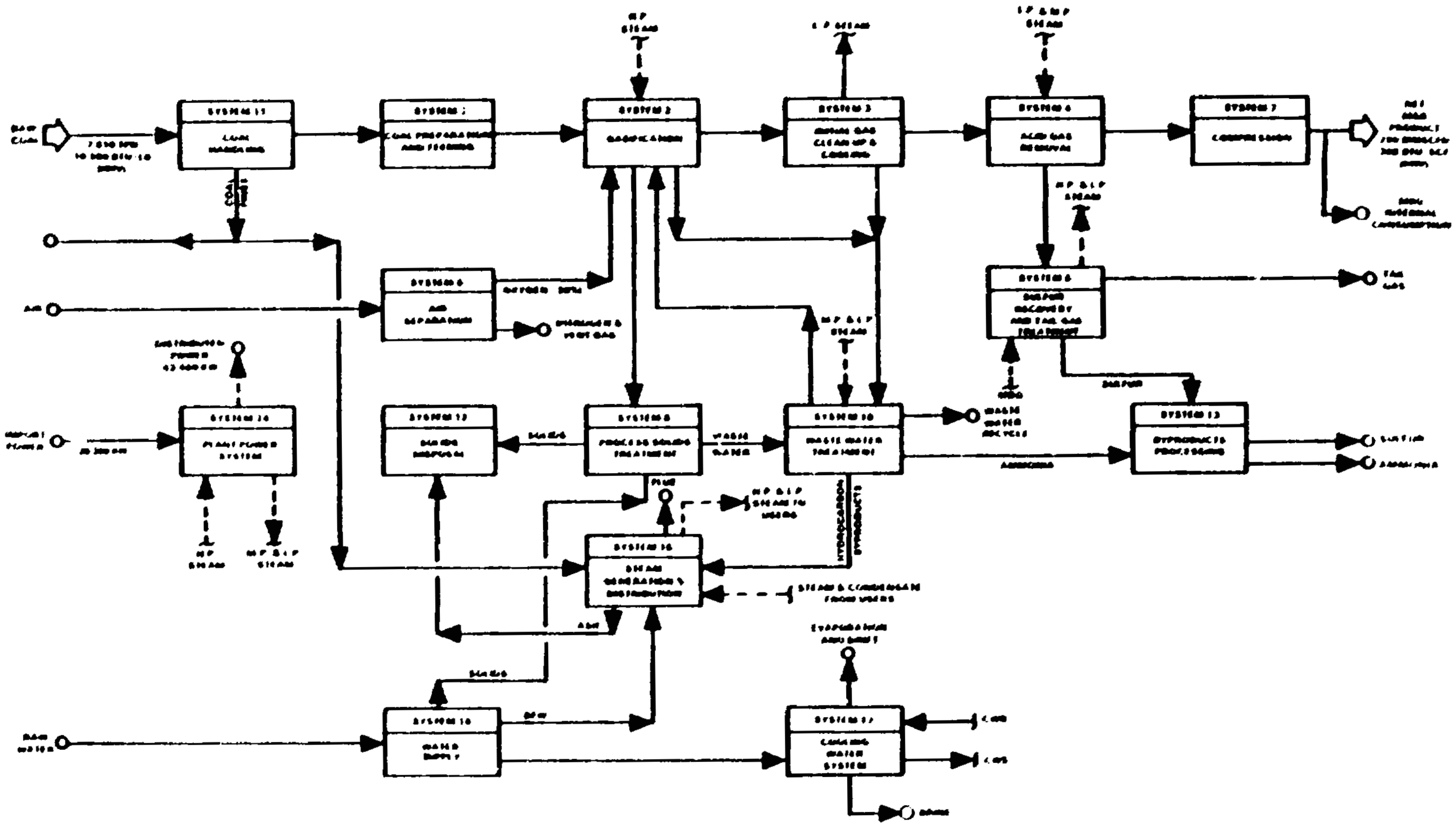


Figure II.D.1. Block Flow Diagram for Lurgi MBG Module

ITEM NUMBER	RAW MATERIAL CHEMICAL OR COMPONENT	ASSOCIATED SYSTEMS	QUANTITY REQUIRED INITIAL	REPLACEMENT	COST	SOURCE	COMMENTS
1	ACTIVATED ALUMINA	98		6300 LB 3 YRS	$407.45 DRUM	PALL TRINITY	FOR OCALA FL 350 LB DRUM
2	ACTIVATED CARBON	05	156 TON	18 TON YR	$1500 TON	ICI AMERICA	FOR MARSHALL TEX 1200 LB PALLET
3	ALGICIDES	19	55 GAL	55 GAL 6 MO	$1004 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUM
4	ALKALINITY CONTROL CHEMICALS	01	400 LB	400 LB MO	$68 DRUM	WATCON INC	FOR SOUTH BEND IND 100 LB DRUM WATCON 1469
5	ALUM	05	200 LB	200 LB MO	$16.81 100 LB	JONES CHEMICAL	FOR MOBILE AL 100 LB BAG
6	AMMONIA	22	1000 LB	1000 LB YR	$84 CYL	INDUSTRIAL CHEMICALS INC	FOR DECATUR AL 150 LB CYL
7	ANTHRA QUINONE DISULFONIC ACID	22				RALPH M PARSONS	QUANTITY COST AND SOURCE PROVIDED BY CLEMSON
8	ANTIFOAM AGENT	04	55 GAL	55 GAL YR	$640 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUM
9	BEAVON CATALYST	16	6 TONS	3 TONS YR	$7000 TON	CATALCO INC	FOR PACIFIC COAST
10	BENFIELD ACTIVATOR	22				ALLIED CHEMICAL	QUANTITY COST AND SOURCE PROVIDED BY CLEMSON
11	CHLORINE	19 05	2 TONS	6 TONS YR	$420 CYL	INDUSTRIAL CHEMICALS INC	FOR DECATUR AL 1 TON CYL
12	CHROMATE INHIBITOR	19	300 GAL	150 GAL 3 MO	$521 DRUM	NALCO CHEMICAL CO	FOR ATLANTA GA 30 GAL DRUM
13	CLAUS CATALYST	16	6 TON	3 TON YR	$1300 TON	KAISER INC	FOR PACIFIC COAST 5 TON MIN
14	DEWATERING AID	11	225 GAL	225 GAL 6 MO	$400 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUM
15	DIMETHYL ETHER OF POLYETHYLENE GLYCOL	22				ALLIED CHEMICAL	QUANTITY COST AND SOURCE PROVIDED BY CLEMSON
16	ETHYLENE DIAMINE TETRA ACETIC ACID (EDTA)	22 04	225 GAL	225 GAL MO	$570 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUMS
17	ETHYLENE GLYCOL	08	3000 GAL	400 GAL YR	.228 DRUM	MOORELAND McKESSON	DELIVERED 519 LB IN A 55 GAL DRUM
18	FERRIC CHLORIDE	05	3100 LB	1250 LB MO	$5.10 DRUM	INDUSTRIAL CHEMICALS INC	FOR DECATUR AL 575 LB DRUM
19	FILTER SAND	05	225 TON	2.25 TON YR	$23.70 TON BAG $17.40 TON BULK	ECONOMY CORP GRAVER WATER DIV	FOR UNION N J 100 LB BAG
20	FREON 12	22	8500 LB	4700 LB YR	$160 CYL	NATIONAL AIR CONDITIONING & REF SUP CO	FOR HUNTSVILLE AL 145 LB CYL
21	HYDRAZINE	04	12 GAL	10 GAL YR	$500 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 30 GAL DRUM
22	LIME	05	2 TON	1 TON MO	$9.76 100 LB	ASHLAND CHEMICAL CO	DELIVERED 100 LB BAG
23	METHANOL	22	7000 GAL	100 GAL MO	$591 DRUM	MOORELAND McKESSON	DELIVERED 385 LB IN A 56 GAL DRUM
24	MOLECULAR SIEVE AIR SEPARATION	08		4000 LB 3 YRS	$514 DRUM	UNION CARBIDE CORP	FOR HOUSTON TX 275 LB IN A 55 GAL DRUM TYPE 13X
25	MOLECULAR SIEVE INSTRUMENT AIR	08		900 LB 3 YRS	$519 DRUM	UNION CARBIDE CORP	FOR HOUSTON TX 300 LB IN A 55 GAL DRUM TYPE 4A
26	MORPHOLINE	06	55 GAL	55 GAL YR	$799 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUM
27	OCTYLDECYLAMINE	04	15 GAL	15 GAL MO	$155 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 15 GAL DRUM
28	PHOSPHORIC ACID	15 17	19.8 TON	10 TON MO	$450 TON	ASHLAND CHEMICAL CO	75% FOR NASHVILLE TENN BULK
29	POLYELECTROLYTE	05	55 GAL	550 GAL YR	$541 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUM
30	POLYMERS (COAGULANT AIDS)	11	225 GAL	225 GAL 6 MO	$471 DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUM
31	POLYPHOSPHATE	19	600 GAL	400 GAL 3 MO	$347 DRUM	NALCO CHEMICAL CO	FOR ATLANTA GA 55 GAL DRUM
32	POTASSIUM CARBONATE	22	500 LB	50 LB 6 MO	$56.60 BAG	INDUSTRIAL CHEMICALS INC	FOR DECATUR AL 100 LB BAG
33	PROPYLENE	22	500 GAL	50 GAL YR	$1.75 GAL	AIR PRODUCTS & CHEMICALS	FOR HOMETOWN PA 50 LB CYL 5 LB GAL
34	RESIN SODIUM ZEOLITE	05	400 FT	400 FT 6 YRS	$50 FT	NALCO CHEMICAL CO	FOR ATLANTA GA
35	RESIN STRONG ACID CATION	05	200 FT	200 FT 6 YRS	$61 FT	NALCO CHEMICAL CO	FOR ATLANTA GA
36	RESIN STRONG BASE ANION	05	200 FT	200 FT 6 YRS	$110 FT	NALCO CHEMICAL CO	FOR ATLANTA GA
37	RESIN WEAK ACID CATION	05	200 FT	200 FT 6 YRS	$61 FT	NALCO CHEMICAL CO	FOR ATLANTA GA
38	RESIN WEAK BASE ANION	05	200 FT	200 FT 6 YRS	$150 FT	NALCO CHEMICAL CO	FOR ATLANTA GA
39	REVERSE OSMOSIS ELEMENTS	05		12 ELEMENTS 3 YRS	$6,000 ELEMENT	CONTINENTAL WATER COND CO	FOR EL PASO TX [illegible] ELEMENT
40	SCALE INHIBITORS	11	110 GAL	550 GAL YR	$6 GAL	PETROLITE CORP	FOR ST LOUIS MO 55 GAL DRUM DISPENSE ETC
41	SILICA GEL	08	4500 LB	4500 LB 3 YRS	$1,777 BAG $247 DRUM	PALL TRINITY	FOR OCALA FL 1000 LB BAG 100 LB DRUM
42	SODIUM BICARBONATE	22	500 LB	100 LB YR	$26.50 100 LB	ASHLAND CHEMICAL CO	DELIVERED 100 LB BAG
43	SODIUM BISULFITE	06	500 LB	500 LB YR	$30.10 100 LB	ASHLAND CHEMICAL CO	DELIVERED 400 LB DRUM
44	SODIUM CARBONATE	22 06	2000-8000 TONS	17,000-30,000 TONS	$10.50 100 LB	ASHLAND CHEMICAL CO	DELIVERED 100 LB BAG
45	SODIUM CHLORIDE	05	2750 LB	2750 LB MO	$412.50 DRUM	INDUSTRIAL CHEMICALS INC	DELIVERED 275 LB DRUM
46	SODIUM HYDROXIDE	10 11 15 17 05	225 TON	24.5 TON MO	$175 TON	ASHLAND CHEMICAL CO	RAYON GRADE 50% FOR MUSCLE SHOALS AL BULK
47	SODIUM SULFITE	04	500 LB	100 LB MO	$18.05 100 LB	ASHLAND CHEMICAL CO	ANHYDROUS 95% DELIVERED 100 LB BAG
48	SULFURIC ACID	10 11 17 19 05	76.5 TON	45.9 TON MO	$49.95 TON	ASHLAND CHEMICAL CO	93% FOR COPPERHILL TENN BULK
49	THICKENING CHEMICALS	11	2000 LB	2000 LB MO	$222 DRUM	HERCULES INC	FOR HOPEWELL VA HERCULES 472 LB IN 55 G DRUM
50	TRIETHYLENE GLYCOL	23	600 GAL	60 GAL MO	$300 DRUM	MOORELAND McKESSON	DELIVERED 522 LB IN A 55 GAL DRUM
51	ULTRA FILTRATION ELEMENTS	05		126 ELEMENTS 3 YRS	$1,300 ELEMENT	CONTINENTAL WATER COND CO	FOR EL PASO TX [illegible] ELEMENT
52	VANADIUM PENTOXIDE	22				RALPH M PARSONS CO	QUANTITY COST AND SOURCE PROVIDED BY CLEMSON
53	ZINC INHIBITOR	19	600 GAL	150 GAL 3 MO	[illegible] DRUM	*NALCO CHEMICAL CO	FOR ATLANTA GA 30 GAL DRUM

*BETZ IS AN ALTERNATE SUPPLIER

Figure II.E.1. Raw Materials Summary

It should be noted that some items in the raw materials list (e.g., Strong Acid Cation Resin) represent a general class of materials and, of necessity, the cost listed represents a specific material in that class. The material chosen for such cases was the one that is often used in applications such as this and has a representative cost. However, the specific material selection must await final system design. The initial and replacement quantities are approximate amounts based upon information from reference literature and vendors and upon prior design experience and engineering judgment. The exact quantities required will be obtained during final system design. The unit cost and transportation information was obtained from the listed commercial sources and represents current market conditions.

The data presented, while not exact, are sufficiently accurate to be used as a basis for determining the operating costs of reference facilities.

F. COAL GASIFICATION BY-PRODUCTS MARKET ANALYSIS

This section describes a market analysis of the by-products from the TVA Coal Gasification Facility located in Northern Alabama. The generic price, use, and future market expectations are shown in Figure II.F.1. for the by-products considered in this analysis. Oxygen is a possible export because all modules will not become operational at the same time and all the oxygen produced will not be consumed until the entire facility becomes operational. However, because of the short-term nature of this situation, oxygen from this plant must be considered "merchant" oxygen rather than tied to a long-term contract. Oxygen, nitrogen, argon, and carbon dioxide are treated as gases, and transportation by pipeline or truck is not considered. Tar, slag, and ash contents are not defined, therefore, they are treated generically. Sulfur is an excellent by-product candidate for marketing.

TVA design criteria specify only truck and barge transportation, so rail costs are not included.

BY-PRODUCT	U.S. POTENTIAL USE	CURRENT PRICE (EARLY 1980)	DISTANCE FROM PLANT WHERE TRANSPORTATION COSTS=CURRENT PRICE (MILES)	TVA MARKET PENETRATION RATING
SLAG/ASH	ROAD AND CEMENT FILL, INSULATION, FLUE GAS DE-SULFURIZATION, LAND FILL	NONE (80% DUMPED)	------	POOR
SULFUR	PHOSPHATATE FERTILIZERS, PAPER, SOIL NUTRIENT, ROAD BASE	\$116/LONG TON + \$30 (SPOT)	> 1500	GOOD
OXYGEN	METAL MANUFACTURING, HEALTH SERVICES, METAL FABRICATING	34¢/100 FT^3 & SHIPPING & EQUIPMENT	------	FAIR
NITROGEN	BLANKETING ATMOSPHERES-CHEMICALS, ELECTRONICS, METALS, FREEZING AGENT, AEROSPACE	30.5¢/100 FT^3 & SHIPPING & EQUIPMENT	------	FAIR
CARBON DIOXIDE	FOOD REFRIGERATION INDUSTRIAL REFRIGERATION, CARBONATION	\$60/TON	1100	GOOD
TAR	CHEMICALS, FUEL	\$95/TON	> 1500	GOOD
STEAM	PROCESS HEAT	\$5-\$6/MMBTU	< 10	FAIR
ARGON	LAMP FILLER, WELDING CUTTING, CRYSTAL GROWTH	73.2¢/100 FT^3	------	FAIR

Figure II.F.1. By-Product Summary

Conclusions of by-product market analysis are:

- Gases will probably have to be purified for market sales.
- Gases will have to liquefied or transported by pipeline over relatively short distances (liquefaction costs have not been determined).
- Most of the by-products will probably be used by new industries locating near the Murphy Hill site.
- Slag and ash will be impounded unless environmental constraints on utilization are relaxed.
- Tar can be converted by processing to fuel for industries during natural gas curtailment. This may be expensive and may lower the price that purchasers are willing to pay for the tar.
- Sulfur is a prime by-product candidate for marketing.

CHAPTER III

TRADE STUDIES

The Reference Facility Design configurations were arrived at by a process of alternate systems evaluations. The systems considered are detailed in "Coal Gasification Systems Engineering and Analysis Task 5.0 - Coal Gasification Catalog". Selection of the final plant configuration was accomplished by a series of trade studies involving the processes available to satisfy each system requirement. Trade studies involved quantitative engineering studies, team experience, and engineering judgment.

In addition to the plant system trade studies, an evaluation of the effects of cleaning coal was also completed. Table III.1 lists the major trade studies performed. Table III.2 lists the options considered in each trade. Based on these studies it is concluded that alternate choices in all cases have less than a 5 percent impact on product price as long as the choices are all technically viable. These trades are discussed in the following sections.

A. Coal Variation Trade Study

Coal variation was studied in order to evaluate the potential value of cleaning coal at the mine site prior to shipping to the gasification plant site. For purposes of this study the plant site was taken as Murphy Hill, Alabama. The study involved determining the cost effects of delivered equivalent MAF coal as a function of degree of cleaning, plant capital, and operating cost. Both conventional washing and deep clean froth flotation were considered.

1. Cost of Delivered Coal

Coal Properties

Three coals are of interest to TVA for use in their gasification project-Kentucky No. 9, Illinois No. 6, and a northern Alabama coal. The Alabama coal would be trucked to the site, supplying about 5% of the gasifier feed.

The properties of these coals are shown in Tables III.A.1, III.A.2 and III.A.3. Data for the Kentucky No. 9 coal were taken from the TVA design specifications (county unidentified). Data for the other two coals are representative of Illinois No. 6, Christian County, Illinois and Mary Lee Seam, Jefferson County, Alabama.

The coal properties are shown for three levels of preparation - Run-of-mine (no preparation), washed, and deep cleaned. The washed coal properties are representative of dense media cleaning of the +28 mesh coal, with the 28mx 0 coal bypassing the cleaning plant but being mixed with the cleaned coal before shipment. The coal properties for this level of washing were estimated from TVA washability data (Ky. No. 9 coal only), U.S. Bureau of Mines publications, Keystone Coal Manual, and other data available to the BDM/Mittelhauser team.

TABLE III.1
MAJOR TRADES

- PROJECT TRADES

  COAL VARIATION TRADES
  ALTERNATE UTILITIES OPTIONS

- SYSTEM CONFIGURATION TRADES

  COMPRESSION
  COAL PREPARATION & HANDLING

- SYSTEM SELECTION TRADES

ACID GAS REMOVAL	PHENOL RECOVERY
COAL FEEDING (TEXACO)	AMMONIA RECOVERY
SULFUR RECOVERY	TAR/OIL DISPOSITION
WATER TREATMENT	BIOLOGICAL OXIDATION
AIR SEPARATION	

CONCLUSIONS

- ALTERNATIVE CHOICE WOULD YIELD LESS THAN 5% INCREASE IN PRODUCT COST
- CHOSEN ALTERNATIVE IS TECHNICALLY AND ECONOMICALLY VIABLE

TABLE III.2
TRADE OPTIONS

TRADE	TYPE	ALTERNATIVES CONSIDERED
COAL RECEIVING & STORAGE	CONFIGURATION	• 4 x 5000 TPD MODULAR SYSTEMS • 1 x 20,000 TPD MODULAR SYSTEMS
ACID GAS REMOVAL	SELECTION	• SELEXOL • RECTISOL • BENFIELD • SULFINOL • STRETFORD
GAS COMPRESSION	CONFIGURATION	• AGR AFTER COMPRESSION • AGR BEFORE COMPRESSION • AGR BETWEEN COMPRESSION STAGES
BYPRODUCT STORAGE & LOADING (LURGI & BGC)	TAR/OIL DISPOSITION	• BURN IN FIRED EQUIPMENT • SELL AS BYPRODUCT
PHENOL RECOVERY	SELECTION	• NON-RECOVERY • PHENOSOLVAN • CHEM-PRO
NH_3 RECOVERY	SELECTION	• NON-RECOVERY • CHEVRON-WWT • PHOSAM-W
SULFUR RECOVERY	SELECTION	• CLAUS + SCOT • CLAUS + BEAVON • CLAUS + WELLMAN-LORD

TABLE III.2 (Continued)

TRADE	TYPE	ALTERNATIVES CONSIDERED
STEAM GENERATION	BOILER SELECTION	• MAXIMIZE PURCHASED POWER, NO BOILERS EXCEPT STARTUP BOILER • COAL-FIRED BOILER WITH FGD • MBG-FIRED BOILER
	SUPERHEATER SELECTION	• NO SUPERHEAT, USE SATURATED STEAM IN DRIVERS • COAL-FIRED SUPERHEATER WITH FGD • MBG-FIRED SUPERHEATER
AIR SEPARATION	CONFIGURATION	• MAXIMUM PURITY O_2, GASEOUS PRODUCT • MINIMUM PURITY O_2, GASEOUS PRODUCT • MAXIMUM PURITY O_2, LIQUID PRODUCT • MINIMUM PURITY O_2, LIQUID PRODUCT
GENERAL FACILITY	PROJECT	COAL VARIATION TRADES
	PROJECT	ALTERNATE UTILITIES OPTIONS
BIOLOGICAL OXIDATION	SELECTION	• AIR ACTIVATED SLUDGE • PRESSURE OXYGEN ACTIVATED SLUDGE
WATER TREATMENT	SELECTION	• TREATMENT FOR RIVER DISCHARGE • ZERO-LIQUID DISCHARGE TO RIVER

The deep cleaned coal properties are representative of crushing the entire ROM coal feed to 28 mesh and cleaning this by two-stage froth flotation. Current coal preparation practice is to clean the coal in a coarse a size as possible, so the deep cleaned coal properties should be considered a "theoretical" limit, beyond what is currently practiced.

(This approach to coal cleaning has been suggested, however, as a possibility for future coal preparation plants). Accordingly, data were not available for deep cleaned properties of the subject coals, and these properties were therefore projected from typical froth flotation results. Thus, the results are more speculative than those shown for the washed coal case.

In addition to these three levels of preparation, a fourth level is shown in Table III.A.1 for the Kentucky No. 9 coal, identified as "Washed with fines disposal." These properties would result from dense medium cleaning of the +28 mesh coal, with disposal of the 28mx0 coal (8% of the ROM feed) at the preparation plant.

Figures III.A.1 and III.A.2 show the size distributions for the 3"x0 washed and 28 mesh x0 deep cleaned products.

Coal Transport Costs

The delivered coal prices shown in Tables III.A.1 and III.A.2 for the Kentucky No. 9 and Illinois No. 6 coals include a cost of \$4/ton for barge transportation. The transport cost for the Mary Lee coal (Table III.A.3), being trucked in, is estimated at \$25/ton. For a range of barge transport costs of \$2-6/ton, the delivered costs of Kentucky and Illinois coals are:

	Delivered coal cost \$/10^6 Btu
Kentucky No. 9, ROM	1.159 - 1.341
Kentucky No. 9, Washed	1.347 - 1.519
Kentucky No. 9, Deep cleaned	1.688 - 1.851
Kentucky No. 9, Washed w/fines disposal	1.448 - 1.616
Illinois No. 6, ROM	1.165 - 1.360
Illinois No. 6, Washed	1.438 - 1.619
Illinois No. 6, Deep cleaned	1.693 - 1.857

Table III.A.1
Coal Properties - Ky. No. 9

COAL	Ky. No. 9 ROM	Ky. No. 9 Washed	Ky. No. 9 Deep Cleaned	Ky. No. 9 Washed w/Fines Disposal
TPD	20000	19002	18039	18556
TOTAL MOISTURE, %	9.564	10.0	10.0	9.0
HHV, Btu/lb (dry basis)	12141	12935	13619	13121
HHV, Btu/lb (as rec'd)	10980	11642	12257	11940
ULTIMATE (dry basis)				
C	67.31	72.41	77.20	73.43
H	4.757	5.12	5.46	5.19
N	1.529	1.64	1.75	1.67
O	6.343	6.82	7.27	6.92
S	4.100	2.94	2.00	2.86
Ash	15.83	10.98	6.23	9.84
Cl	0.121	0.09	0.09	0.09
PROXIMATE (dry basis)				
VM	37.54	39.70	41.82	40.21
FC	46.63	49.32	51.95	49.95
Ash	15.83	10.98	6.23	9.84
PYRITIC SULFUR	2.517	1.339	0.30	1.230
ORG + SULFATE SUL	1.583	1.601	1.70	1.630
ASH SOFTENING TEMP (Reducing) ^{o}F	2031	1981	1981	1981
FSI	3.0-6.5	3.0-6.5	3.0-6.5	3.0-6.5
GRINDABILITY INDEX	59	57	57	57
SIZE	8"x0	3"x0	28mx0	3"x28m
Wt. Recovery	-	90.15	83.8	82.15
Btu Recovery	-	96.06	94.0	88.8
\$/ton, FOB (moist)	23.45	29.37	39.37	32.59
\$/ton, delivered (moist)	27.45	33.37	43.37	36.59
$\$/10^6$ Btu, FOB	1.068	1.261	1.606	1.365
$\$/10^6$ Btu, delivered	1.250	1.433	1.769	1.532

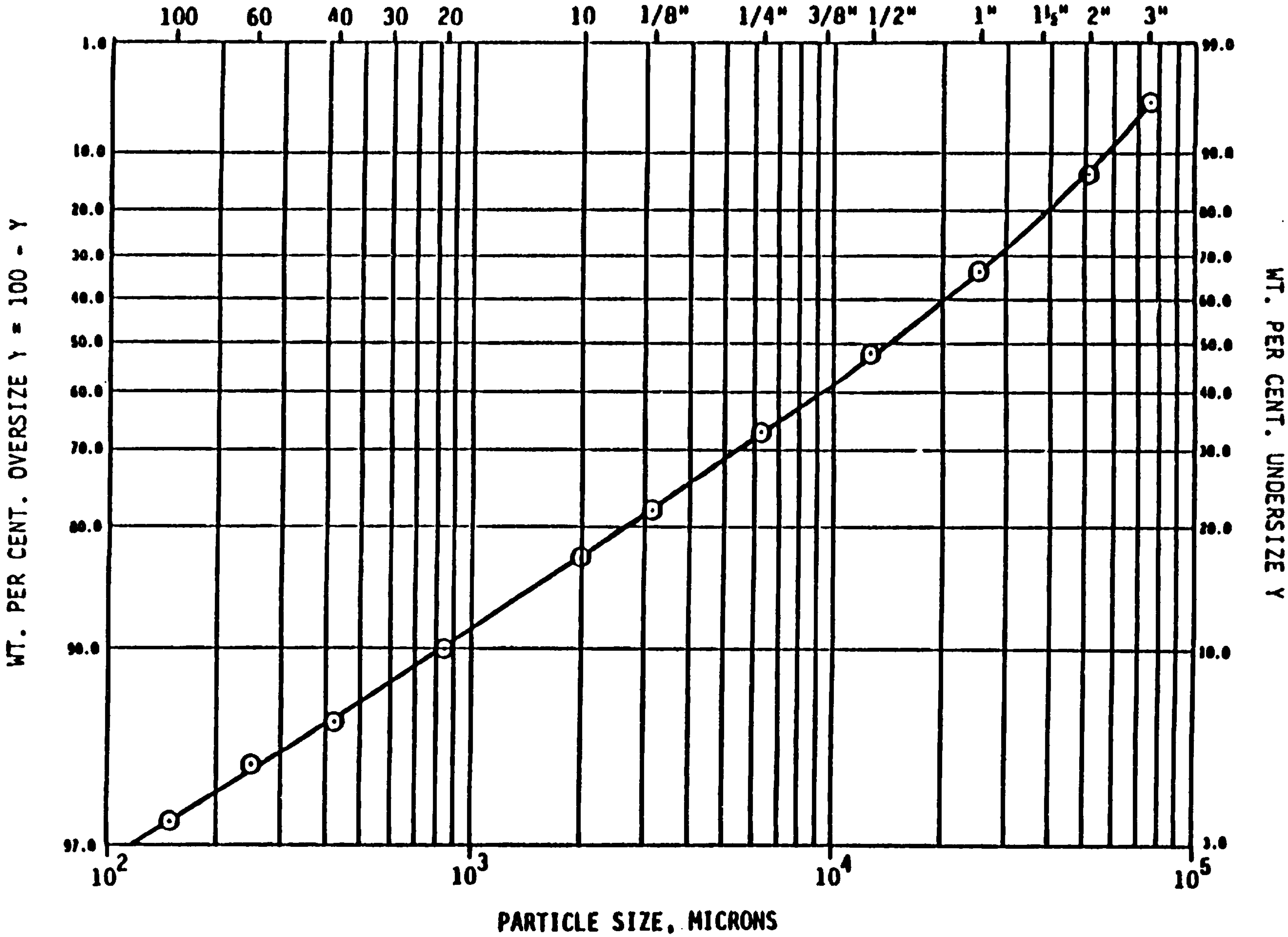
Table III.A.2
Coal Properties - Ill. No. 6

COAL	Ill. No. 6 ROM	Ill. No. 6 Washed	Ill. No. 6 Deep Cleaned
COUNTY	Christian	Christian	Christian
TPD to gasifiers	21682	19943	18220
TOTAL MOISTURE	14.35	14.2	10.0
HHV, Btu/lb (dry basis)	12000	12919	13590
HHV, Btu/lb (as rec'd)	10278	11085	12231
ULTIMATE (dry basis)			
C	62.54	68.68	73.34
H	4.56	5.01	5.35
N	1.15	1.26	1.35
O	8.55	9.39	10.03
S	4.90	4.42	2.56
Ash	18.02	11.03	7.16
Cl	0.28	0.21	0.21
PROXIMATE (dry basis)			
VM	39.71	42.97	44.97
FC	42.27	46.00	47.87
Ash	18.02	11.03	7.16
PYRITIC SULFUR	2.85	2.09	0.34
ORG + SULFATE SUL	2.05	2.33	2.22
ASH SOFTENING TEMP (Reducing) ^{o}F	2100	2100	2100
FSI	2-5	2-5	2-5
GRINDABILITY INDEX	56	55	55
SIZE	8"x0	3"x0	28mx0
Wt. Recovery	-	83.6	83.6
Btu Recovery	-	90.0	94.0
\$/ton, FOB (moist)	21.95	29.89	39.42
\$/ton, delivered (moist)	25.95	33.89	43.42
\$/10^6 Btu, FOB	1.068	1.348	1.611
\$/10^6 Btu, delivered	1.262	1.529	1.775

Table III.A.3
Coal Properties - Mary Lee

COAL (NORTHERN ALABAMA)	Mary Lee ROM	Mary Lee Washed	Mary Lee Deep Cleaned
COUNTY	Jefferson	Jefferson	Jefferson
TPD to gasifiers	19143	17261	16295
TOTAL MOISTURE	2.0	2.0	10.0
HHV, Btu/lb (dry basis)	12860	13780	14653
HHV, Btu/lb (as rec'd)	12603	13504	13188
ULTIMATE (dry basis)			
C	69.99	77.8	80.38
H	4.41	4.9	5.06
N	1.53	1.7	1.76
O	4.32	4.8	4.96
S	0.90	0.8	0.30
Ash	18.85	10.0	7.54
Cl	-	-	-
PROXIMATE (dry basis)			
VM	26.35	29.4	30.02
FC	54.80	60.6	62.44
Ash	18.85	10.0	7.54
PYRITIC SULFUR	0.71	0.60	0.09
ORG + SULFATE SUL	0.19	0.20	0.21
ASH SOFTENING TEMP (Reducing) ^{o}F	2700	2810	2810
FSI	8	8.5	8.5
GRINDABILITY INDEX	82	82	82
SIZE	8"x0	3"x0	28mx0
Wt. Recovery	-	87.8	82.5
Btu Recovery	-	94.1	94.0
$/ton, FOB (moist)	26.92	34.56	41.67
$/ton, delivered (moist)	51.92	59.56	66.67
$\$/10^6$ Btu, FOB	1.068	1.280	1.580
$\$/10^6$ Btu, delivered	2.060	2.205	2.528

Figure III.A.1. Size Distribution - 3" x 0 Coal



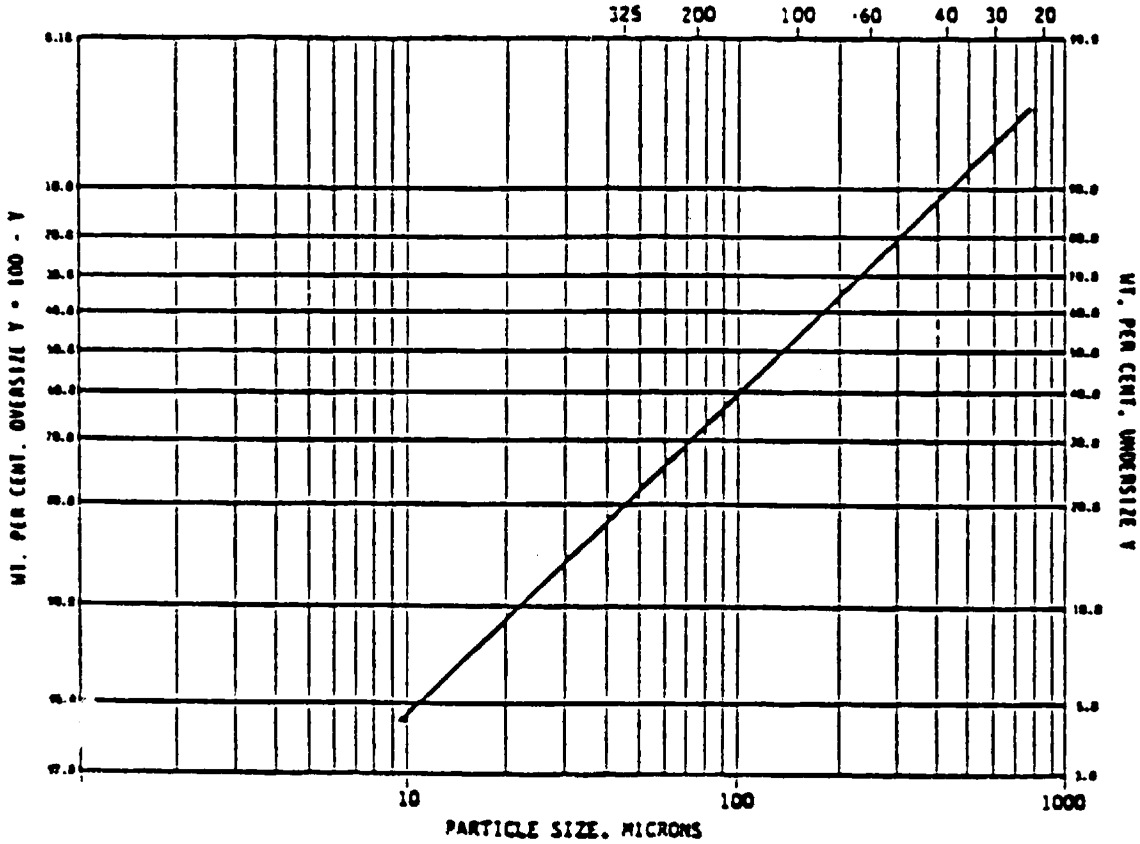


Figure III.A.2. Size Distribution - 28 Mesh x 0

Deep Cleaning Option

As described earlier, the coal properties shown for deep cleaned coal are based on a hypothetical froth flotation plant. It should be possible to deep clean these coals to an ash content of 3-5 percent, as is currently practiced at the Homer City, Pennsylvania coal preparation plant. The Homer City plant produces two products - a deep cleaned coal and a middlings fraction intermediate between the deep cleaned and the reject stream. The quantity of middlings produced is about twice that of the deep cleaned coal. The ash and sulfur contents of the middling fraction are only slightly reduced compared to ROM coal.

Deep cleaning coal in this fashion might be a viable option for the TVA gasifiers, provided that a 40,000 TPD market is available for the middling fraction.

Fine Coal Handling

The delivered costs for deep cleaned coal shown in Tables III.A.1, III.A.2, and III.A.3 include the same transport costs ($4/ton barge, $25/ton truck) as for ROM and washed coal transport. Current coal handling practice, however, does not provide a means for handling large tonnages of 28 mesh x 0 coal. While such coal might be transported in covered barges, conventional coal handling equipment is not suitable for loading and unloading these carriers without serious dust losses.

In addition, long term storage at the gasifier site would have to be enclosed. In order to provide 90 days of dead storage, as required in the TVA specifications, the cost of concrete silos alone would be approximately $230 million.

Alternatively, a realistic comparison of delivered deep cleaned coal costs to the ROM and washed coal cases can be made by adding the cost of coal agglomeration to the cost of the deep cleaned coal. The cost of agglomeration is estimated at $6.93/ton coal.

The cost effects at the plant site are given in Tables III.A.4 and III.A.5.

Coal Prep Plant Bypass

One possibility of savings is, that after specifying a maximum ash content from a coal supplier, allowing the coal supplier to bypass cleaning of the coal that already met the specification. Using the Kentucky No. 9 coal as an example, with an 11% ash specification, 5.9% of the ROM coal would meet the ash specification without cleaning. This amount of coal is not large enough to warrant the intensive sampling that would be required to overcome the practical problem of identifying

Table III.A.4
SYSTEM 11
COAL RECEIVING & STORAGE
CAPITAL COST, JAN. 1980 $ (MM$)

DESCRIPTION	Ky. No. 9 ROM*	Ky. No. 9 Washed	Ill. No. 6 ROM	Ill. No. 6 Washed
1. Barge Unloading & Material transfer equipment	10.0	9.8	10.4	10.0
2. Open coal storage piles, Stacking/reclaiming	30.0	29.2	31.2	30.0
3. Rotary breakers	1.0	-	1.0	-
4. Concrete silos	6.0	5.7	6.5	6.0
5. Truck dump hopper	0.2	0.2	0.2	0.2
6. Conveyors not included above				
a. Truck station to crusher	0.3	-	0.3	-
b. Crusher to silos	0.6	-	0.6	-
c. Truck station to silos	-	0.3	-	0.3
SUBTOTAL	48.1	45.2	50.2	46.5
INDIRECT COSTS	16.1	15.2	17.1	15.8
TOTAL	64.2	60.4	67.3	62.3

* Base Case

Table III.A.5
SYSTEM 11
OPERATING REQUIREMENTS

	Ky. No. 9 ROM*	Ky. No. 9 Washed	Ill. No. 6 ROM	Ill. No. 6 Washed
ELECTRICAL - Million kwh/yr				
Barge unloading	1.64	1.56	1.78	1.64
Rotary breakers	0.27	-	0.29	-
Conveyor to silos	1.19	1.13	1.29	1.19
Conveyor from truck unloading**	-	-	-	-
Stockpiling conveyors & Stacker**	-	-	-	-
Other conveyors	1.25	1.19	1.36	1.25
	4.35	3.88	4.72	4.08
OIL - 1000 gal/yr				
Mobile equipment	103.4	98.2	112.1	103.1

* Base case

** Assumes dead storage pile is inactive.

and segregating the actual coal that could be bypassed. We conclude that no savings are likely from taking advantage of variability in coal ash content.

2. Plant Cost Savings

For a 20,000 TPD plant the use of washed vs. ROM Kentucky No. 9 coal impacts the annual coal cost as follows:

	ROM CASE	WASHED CASE	DELTA
Annual coal cost - MM$/YR	180.35	204.82	24.47

The increase in annual coal cost due to the use of washed coal is compensated somewhat by reductions in the capital investment for the following systems:

	TOTAL FACILITY CAPITAL INVESTMENT - MM$		
	ROM CASE	WASHED CASE	DELTA
Coal Handling	64.214	60.369	-3.845
Final Solids Disposal	85.976	69.840	-16.136
Process Solids Treatment	3.440	2.868	-0.572
Sulfur Recovery	65.928	51.315	-14.613
			-35.166

These reductions are due to the decrease in the amounts of ash and sulfur contained in the coal as a result of washing. Little or no reduction in the capital investment for the acid gas removal system is expected. This is based on the assumption that the same solvent circulation rate is maintained to provide the same degree of CO_2 removal for the two cases.

Based on the computed changes in costs of coal and capital investment and using a method which is based on factors derived from the detailed procedure of the ROM case, the 1980 cost of product is estimated to be as follc :

	ROM CASE	WASHED CASE	DEEP CLEAN
$/MM Btu	4.82	5.00	5.34

Operation and maintenance effects from cleaning coal have not been analyzed in detail. However, based on percentage of capital cost reduction the effect would be small, not exceeding 10-15 cents per MM Btu as shown by small changes in Table III.A.5.

Therefore, it is concluded that mine mouth coal cleaning is not justified on cost alone. No consideration of the preference for disposing of ash away from the plant has been considered here.

For a design using water slurry feed, such as Texaco, additional savings are realized in the oxygen plant. This is due to the decrease in heat requirement for melting ash and vaporizing water. Associated capital savings are estimated to be 6.48 MM dollars on a plant basis.

3. Coal Variability Evaluations

Coal characteristics vary within a given seam and if seams are blended this variability may be appreciable. This design accommodates up to 2 deviations from the Kentucky No. 9 coal design characteristics for sulfur and ash by:

- Design contingencies typical to coal conversion systems
- Higher frequency of intermittent operations
- Excess capability in acid gas removal
- Extra 25% sulfur recovery capacity for facility

Switching to a coal different from the design coal presents additional problems. The use of only Illinois #6 ROM coal may possibly require the addition of minor design margins to accommodate higher concentrations of ash and sulfur. Blending coal requires special attention to variability in the blend as feed to the gasifiers. The differences in the ash and sulfur of the North Alabama coal are greatly diluted if it comprises only 5 percent of the feed. It would have a minor (5 percent) reduction in total sulfur and its high ash fusion temperature may have a minor impact on ash overheat.

B. Alternate Utilities Options

In the design of the coal gasification facility considerable power is required especially in the Air Separation System and Product Compression System. Studies indicate that where available process derived steam with appropriate superheating should be used for this purpose.

For power requirements over and above available process steam there exists an option of generating steam or purchasing electricity. This trade study defines the break-even cost of electricity as a substitute for facility generated steam and turbine power system. In this study both coal fired boiler and MBG fired boiler steam are considered. Basic assumptions used are given in Table III.B.1, Table III.B.2 and Table III.B.3. Steam rates are given in Table III.B.4, III.B.5 and III.B.6. Figure III.B.1 gives capital costs for the various motors and turbines as a function of horsepower.

Annual operating costs are given by:

Annual Operating Costs (electricity) = C_{AE} =

$$HP \times 0.7457\ KW/HP/eff \times (7884\ HR/YR \times C_e + C_{Fe})$$

where HP = Horsepower

KW = Kilowatts

C_e = Cost of electricity - dollars per KW - hr (use charge)

C_{Fe} = Fixed electrical costs - Dollars/YR-KW (demand charge)

Annual Operating Costs (steam and cooling water) = C_W

$$= HP \times lb\ steam/HP\ hr \times 7884\ HR/YR \times C_{ST}$$

$$+ \frac{(On\ duty,\ BTU/HP\ hr)\ (HP)\ (7884\ HR/YR)}{(1\ \frac{BTU}{lb^{o}F})\ (115^{o} - 85^{o}F)\ (8.34\ lb/gal)} \times C_{CW}$$

where C_{CW} = cost of cooling water dollars per gallon

C_{ST} = cost of steam dollars per pound.

For purposes of this study costs have been assumed as follows:

C_e = \$0.01747/kwhr (year 1980)

C_{Fe} = \$60.00/yr-kw

C_{ST} = \$10.48/Mlbs from MBG Boiler

= \$4.69/Mlbs from MBG Boiler with process use of exhaust steam

C_{ST} = \$5.44/Mlbs from coal fired boiler.

Annual cost of service for power is given by

$$A_{EM} = 0.13388\ P_{EM} + 0.01\ P_{EM} + C_{AE}$$

$$A_{CT} = 0.13388(1.1)\ P_{CT} + .04\ P_{CT} + C_{SW}$$

$$A_{BT} = 0.13388\ P_{BT} + .04\ P_{BT} + C_{S}$$

where:

A_{EM}, A_{CT}, A_{BP} = Annual cost of service for electric motors, condensing turbines, back pressure turbines

P_{EM}, P_{CT}, P_{BT} = Cost of electric motors, condensing turbines, back pressure turbines.

Annual maintenance cost = 0.01 P_{EM} for electric motors

= 0.04 P_{CT} for condensing turbines

= 0.04 P_{ET} for back pressure turbines

Operating life = 20 years
Cost of money = 12 percent

Table III.B.7 gives results of selected examples comparing the use of these alternatives.

CONCLUSION

In addition to the comparisons based on projected costs, the break even values for the purchase of electricity have been determined for the case of 6000 HP drivers operating on 1450 psig 900°F steam produced from two 75 percent boilers. Displacing $10.48/Mlb steam electricity must cost less than $0.085/Kwh. Displacing $5.44/Mlb steam electricity must cost less than $0.041/Kwh.

TABLE III.B.1
UTILITY OPTION ASSUMPTIONS

ELECTRIC MOTORS;	• Annual maintenance equals 1% purchase price • Will last 20 years w/o replacement • Optimum motor type is selected (1st cost vs eff.) (e.g. less than 500 HP are squirrel case induction; greater than 500 HP are conventional synchronous) • Electric motors applied directly to liquid handling driven equipment & with gear unit for gas handling driven equipment • Annual capital charges equal 0.13388 x costs.
STEAM TURBINES;	• Annual maintenance equals 4% purchase price • Turbines will last 20 years w/o replacement • Condensers will be replaced at 10 years • All turbines are 5000-6000 RPM and efficiency floats with horsepower class • Condensing will be to 4" Hg Abs based on average condenser service and Std. cooling water • Up to 1000 HP, single state (backpress only) turbines will be employed • Above 1000 HP, multistage turbines are used • Turbines are applied directly to gas handling driven equipment and require gear unit when applied to liquid handling equipment • Annual capital charges equal 0.13388 x cost.

TABLE III.B.2

MULTISTAGE TURBINE EFFICIENCIES (1%)*

H.P.	1450 B.P.	1450 COND.	650 B.P.	650 COND.
1000	59.0	63.0	57.0	62.0
2000	62.0	66.0	60.0	65.0
3000	64.5	68.5	63.0	68.0
4000	66.5	70.5	65.5	70.5
5000	68.5	72.5	67.5	72.5
6000	70.0	74.0	69.0	74.0
7000	71.0	75.0	70.0	75.0
8000	72.0	76.0	71.0	76.0
9000	72.5	76.5	71.5	76.5
10,000	73.0	77.0	72.0	77.0

* Inlet steam conditions - 1450 psig at 900^{O}F, and 650 psig at 750^{O}F.

TABLE III.B.3

ELECTRIC MOTORS

Utility Annual Operating Cost (C_{AE})

H.P.	Eff. %	$/YR	$/HP-YR
100	91.	16,006.	160.1
200	91.3	31,908.	159.5
300	91.7	47,653.	158.8
400	92.	63,330.	158.3
500	92.3	78,906.	157.8
600	92.7	94,278.	157.1
700	93.	109,637.	156.6
800	93.3	124,896.	156.1
900	93.7	139,908.	155.5
1000	94.	154,958.	155.0
2000	94.3	308.930.	154.5
3000	94.7	461,437.	153.8
4000	95.	613,306.	153.3
5000	95.3	764,219.	152.8
6000	95.7	913,230.	152.2
7000	96.	1,062,106.	151.7
8000	96.3	1,210,054.	151.3
9000	96.7	1,355,679.	150.6
10000	97.	1,501,652.	150.2

1st yr. annual operating cost: (utility)
90% on stream, 100% load, 8760 HRS/YR.

$$= \frac{(\text{H.P.}) \times (0.7457\ \text{KW/HP})}{\text{eff.}} \left[(8760. \times .9)\ \frac{\text{HRS}}{\text{YR.}} \times (\$0.01747/\text{KWhr}) + (\$4.80)(12.) \right]$$

$$= \frac{\text{H.P.}}{\text{eff.}} \times (145.6602)$$

TABLE III.B.4

CONDENSING TURBINE STEAM RATES* AND DATA

H.P.	1450 psi lbn/HP-HR	% Moisture	Cond. Duty ($\frac{BTU}{HP-HR}$)	650 psi lbn/HP-HR	% Moisture	Cond. Duty ($\frac{BTU}{HP-HR}$)
1000.	7.72	1.4	7951.6	9.14	1.7	9414.4
2000.	7.36	2.9	7581.6	8.72	3.0	8982.7
3000.	7.10	4.3	7314.5	8.34	4.3	8592.0
4000.	6.89	5.3	7098.7	8.04	5.4	8283.6
5000.	6.70	6.3	6903.5	7.82	6.3	8057.5
6000.	6.57	7.1	6769.9	7.66	7.0	7893.0
7000.	6.48	7.6	6677.4	7.56	7.3	7790.1
8000.	6.40	8.1	6595.2	7.46	7.8	7687.4
9000.	6.35	8.4	6543.8	7.41	8.0	7635.9
10000.	6.31	8.6	6502.7	7.36	8.25	7584.6

$$* \text{ lbm steam/HP-HR} = \frac{2545 \text{ BTU/HP-HR}}{(h_1 - h_2)\ \frac{\text{BTU}}{\text{lb steam}} \times \text{eff}}$$

TABLE III.B.5

SINGLE STAGE BACKPRESSURE TURBINE STEAM RATES (lb/HP-HR)

HP	Eff. (%)	1450 psig/150 psig*	650 psig/150 psig*
100	31.	35.68	46.51
200	32.	34.56	45.06
300	33.	33.52	43.69
400	34.	32.53	42.41
500	35.	31.60	41.20
600	36.	30.72	40.05
700	37.	29.89	38.97
800	38.	29.11	37.95
900	39.	28.36	36.97
1000	40.	27.65	36.05

* Supply pressure/backpressure

TABLE III.B.6

MULTI-STAGE BACKPRESSURE TURBINE STEAM RATES (lb/HP-HR)

HP	1450 psig/150 psig*	650 psig/150 psig*
1000	18.75	25.30
2000	17.84	24.03
3000	17.15	22.89
4000	16.63	22.01
5000	16.15	21.36
6000	15.80	20.90
7000	15.58	20.60
8000	15.36	20.31
9000	15.26	20.17
10000	15.15	20.03

* Supply pressure/backpressure.

TABLE III.B.7
ALTERNATE UTILITIES
EVALUATION & DECISION

EXAMPLES OF ANNUAL COST OF SERVICE, $\$10^3$

BHP	100	500	1000	5000	10,000
DRIVER TYPE					
ELECTRIC MOTOR (TEFC, GEAR)	16.9	84.0	163.9	805.2	1582.2
BACKPRESSURE TURBINE (WITH GEAR)					
650 PSIG/750°F/150 psig	161.1	711.2	873.8	3701.6	6923.4
1450 PSIG/900°F/150 psig	123.9	546.2	648.7	2806.1	5245.9
CONDENSING TURBINE (WITH GEAR)					
650 PSIG/750°F/4"HgA	--	--	434.7	1798.5	3349.1
1450 PSIG/900°F/4"HgA	--	--	371.5	1549.2	2881.8

o NOTES FOR THE DERIVATION OF THE EXAMPLE OF ANNUAL COST OF SERVICE

- STEAM COST OF $\$5.44/10^3$ LB BASED ON COAL FIRED BOILERS (2 x 70%) WITH FGD AT A PRODUCTION RANGE OF 700,000 TO 800,000 LB/HR AND COAL AT $1.25/MM MTB
- BACKPRESSURE TURBINE STEAM CREDIT AT $1.25/MM BTU ($\$1.08/10^3$ LB)
- 1980 PURCHASE PRICE FOR ELECTRIC POWER EQUAL $0.0174/KW HV.

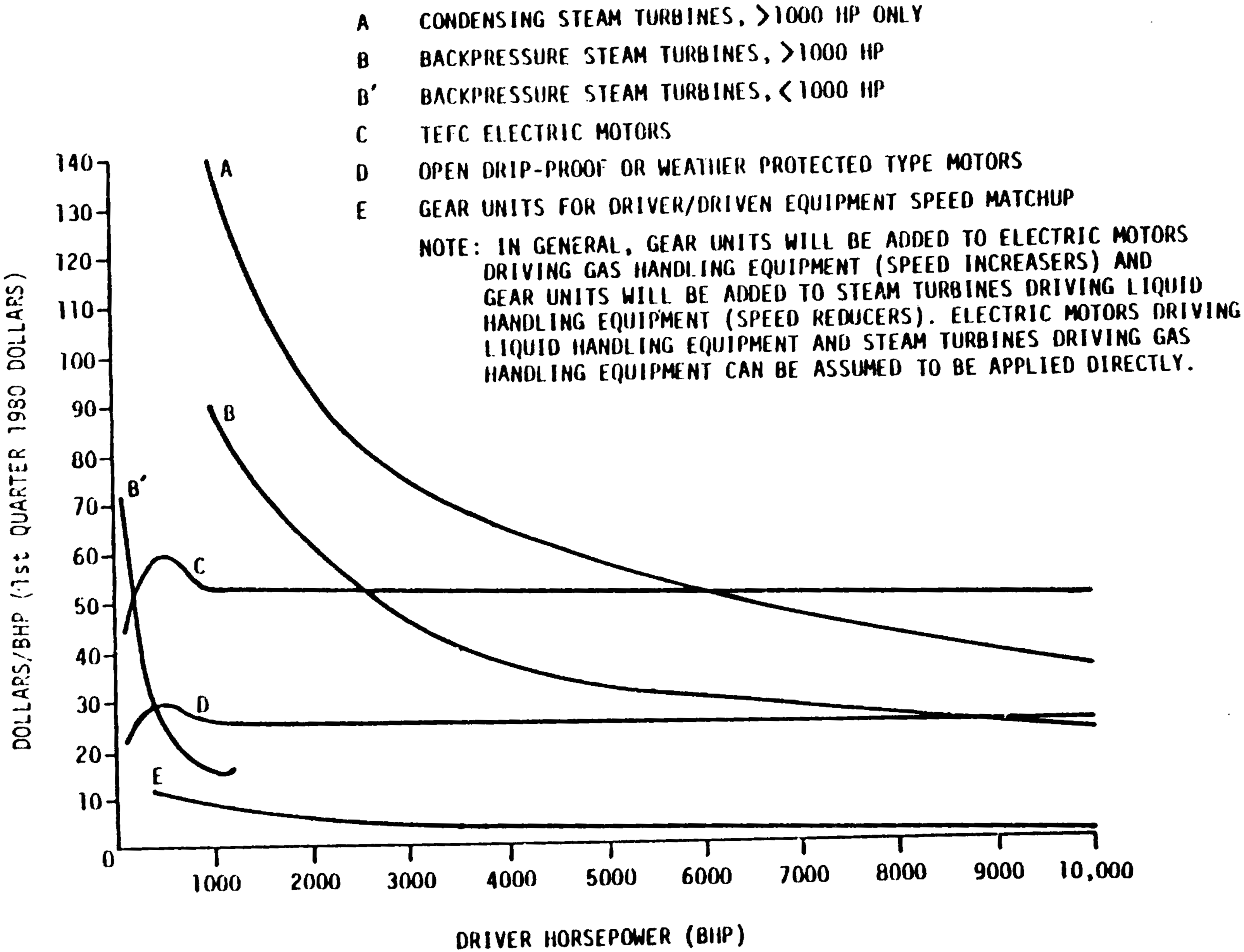


Figure III.B.1. Driver Horsepower Vs. Dollars/BHP

C. Boiler Fuel Trades

The Alternate Utility Trade study reported in Section B required that a basis be set for the production of steam in the plant. Therefore, even though it was decided that no boilers would be included in the Plant, a trade study was performed to compare the production of steam using product MBG and coal as fuel.

The study consisted of evaluating the capital and O&M costs for construction and operation of steam plants at three size levels for both MBG and coal fired boilers. Each plant was assumed to contain 2-75 percent capacity units. Capital costs for the MBG boilers were taken from Guthrie. Coal fired boiler costs were taken from the EPA report 600/7-79-178a. All cost were adjusted to a 1980 basis. Table III.C.1 contains the capital costs for the six cases studied. The coal fired boiler case requires stack gas cleaning. The Wellman-Lord process was used in this study. Table III.C.2 gives the capital and operating costs for this system.

TABLE III.C.1
STEAM GENERATION CAPITAL COSTS

STEAM CAPACITY (M lb/YR)	CAPITAL COSTS (M DOLLARS) MBG	COAL*
664	47,540	74,225
500	38,451	62,967
400	32,159	55,076

*Includes Wellman-Lord stack gas cleanup.

TABLE III.C.2
CAPITAL AND OPERATING COSTS (WELLMAN-LORD)

ITEM	CAPACITY (Mlbs/HR) 664	500	400
Soda Ash (T/YR)	1,578	1,056	845
Water (MMG/YR)	1,532	1,019	815
Power (Mwh/YR)	18,125	12,136	9,709
Steam use (Mlb/YR)	156,000	105,000	84,000
Sulfuric Acid Production (T/YR)	32,206	21,571	17,254
Capital Cost ($MM)	21.135	18.186	15.907
*Annual Material Costs ($MM/YR NET)	-0.524	-0.351	-0.281
Other Operation	5.078	3.993	3.492

*Sales of sulfuric acid more than offset other material costs

Prices assumed for this study are as follows:

TABLE III.C.3
COMMODITY PRICE

COMMODITY	PRICE
Coal	$1.12/MMBTU
MBG	$6.92/MMBTU
Soda Ash	$70/Ton
Water	$0.05/1000 gal
Power	$0.0175/Kwh + demand
H_2SO_4 (Sales)	$60/Ton

Cost analyses for the three MBG fired cases are given in Table III.C.4. Cost analyses for the three coal fired cases are given in Table III.C.5.

TABLE III.C.4
MBG BOILER SUMMARY

COST ITEM (ANNUAL)	STEAM CAPACITY (Mlb/HR) 664	500	400
CAPITAL CHARGES ACCF = 0.135*	6,418,000	5,190,000	4,341,000
FUEL @ 6.92	48,071,000	36,198,000	28,958,000
OTHER OP. COST	4,842,000	3,916,000	3,276,000
TOTAL ANNUAL COST	59,331,000	45,304,000	36,575,000
STEAM COST DOLLARS/Mlb	11.33	11.49	11.60

TABLE III.C.5
COAL BOILER SUMMARY

COST ITEM (ANNUAL)	STEAM CAPACITY (Mlb/HR) 664	500	400
CAPITAL CHARGES ACCF = 0.135*	10,020,000	8,501,000	7,435,000
FUEL COST	7,726,000	5,814,000	4,652,000
W-L NET OPERATION	4,554,000	3,642,000	3,211,000
BOILER OPERATION	5,722,000	4,826,000	4,221,490
TOTAL COST	28,022,000	22,783,000	19,519,000
STEAM COST DOLLARS/Mlb	5.35	5.78	6.19

* Annual capital charges including replacement capital allowances.

The analysis given has not considered the cost of ash disposal. The impact of this cost has been estimated based on the following:

- ash in coal - 15.83 percent
- ash density - 50 lbs per cu ft
- ash pile height - 10 feet
- life of plant - 20 years
- cost of land - $3000/acre
- steam capacity - 664,000 lb per hr

The capital cost of the basis is estimated to be 5.9 million dollars. A total of 83 acres are required for the job. Assuming 20 acres of additional land is required, land cost is 311,000 dollars. Additional operating costs are estimated at 1.0 million dollars per year. The final result is an additional 36 cents per pound of steam.

The cost of steam shown here assumes no secondary use such as heat tracing or other process heating service. If exhaust steam from a back-pressure turbine is used for process heating and credited at fuel costs the cost of steam to its primary use (i.e., B.P. turbine) is reduced accordingly. In the case of MBG fired boilers the net cost is reduced to $4.69 per thousand pounds.

D. Coal Handling Trades

Two approaches to the Coal Handling facilities, System 11, were considered for this design. Alternative one is the construction of separate facilities for each module of the plant with the objective of deferring some capital expenditures in accordance with the overall 20,000 TPD plant construction schedule and perhaps to maximize flexibility. Alternative two is to build a common single 20,000 TPD system to service all of the four modules, minimizing the total system investment and land requirements. The specific considerations are as follows:

ALTERNATIVE NO.	1	2
DESCRIPTION	4x5000 TPD MODULES	1x20,000 TPD FACILITY
MEETS PERFORMANCE CRITERIA	Yes- Excess Barge Capacity	Yes
RELATIVE CAPITAL COST	75% Higher	Base
RELATIVE ENERGY CONSUMPTION	Same	Same
PROPRIETARY	No	No
COMMERCIALLY PROVEN	Yes	Yes
APPLIED DEVELOPMENT NEEDS	None	None
COULD DELAY IMPLEMENTATION	Slightly	No
DATA AVAILABILITY	None	Same
SYSTEM-LEVEL DESIGN	-	-
COST	-	-
SUBSYSTEM LEVEL DESIGN	-	-
COST	-	-
RELATIVE COMPLEXITY	More	Base
RELATIVE OPERATING COST	Higher	Base
POTENTIAL ENVIRONMENTAL PROBLEM	Same	Base
BYPRODUCT MARKETABILITY	Same	Same

The conclusion is that alternative two is preferable.

E. Acid Gas Removal Trades

System 4, Acid Gas Removal, has the objective of meeting the specification of 200 ppmv sulfur in the product gas at lowest cost consistent with reliable performance. The processes considered for this system are Selexol, Rectisol, Benfield, Sulfinol, and Stretford. Tables III.E.1 and III.E.2 summarizes the comparisons considered in making the design selection.

Previous studies (FE-2240-49) by the design teams have shown that process attractiveness varies with the acid gas partial pressure in the AGR process feed. At pressures near 600 psig Benfield, Selexol, and Rectisol are preferable. Likewise, compression of the gas prior to acid gas removal has been shown to be perferable in that additional cost of compressing additional gas (acid gas) is more than offset by reductions in the cost of physical solvent. Selective removal of sulfur bearing gases is preferred in that removal of other acid gases is not required to meet the product specifications.

Table III.E.1
ACID GAS REMOVAL TRADES

ALTERNATIVE NO.	1	2	3	4	5
DESCRIPTION	SELEXOL	RECTISOL	BENFIELD	SULFINOL	STRETFORD
MEETS PERFORMANCE CRITERIA	With chilled solvent or COS hydrolysis	Yes	With COS hydrolysis	Yes	With COS hydrolysis
RELATIVE CAPITAL COST	Base	Approx. same	Approx. same	Approx. same	?
RELATIVE ENERGY CONSUMPTION	Base	Higher	Higher	Approx. same	Lower
PROPRIETARY	Yes	Yes	Yes	Yes	Yes
COMMERCIALLY PROVEN	Yes all except COS hydrolysis	Yes	Except COS hydrolysis	Yes	At low pressure
APPLIED DEVELOPMENT NEEDS	COS hydrolysis if used	None	COS hydrolysis formate preven-tion	None	COS hydrolysis purge dispo-sition
COULD DELAY INPLEMENTATION	Not with chilled solvent	No	Yes	No	Yes
DATA AVAILABILITY					
SYSTEM-LEVEL DESIGN	Yes	Not this ap-plication	No	No	Yes
COST	Yes	No	No	No	Yes
SUBSYSTEM LEVEL DESIGN	No	No	No	No	No
COST	No	No	No	No	No
RELATIVE COMPLEXITY	Base	Higher	Higher	Higher	Same
RELATIVE OPERATING COST	Base	Slightly Higher	Slightly Higher	Slightly Higher	Higher
POTENTIAL ENVIRONMENTAL PROBLEM	No	No	Yes, Formates	Yes, Degradation	Yes
BYPRODUCT MARKETABILITY	No problem	No problem	No problem	No problem	Sulfur may need storage disposal

TABLE III.E.2
ACID GAS REMOVAL EVALUATION
& DECISION

EVALUATION:

- PREVIOUS TEAM STUDIES (FE-2240-49) HAVE SHOWN THAT
  - ATTRACTIVENESS OF PROCESSES VARIES WITH PARTIAL PRESSURE OF ACID GAS IN FEED.
  - AT PRESSURES NEAR THOSE OF THIS STUDY (600 PSIG) PERFERENCE IS BENFIELD, SELEXOL, RECTISOL.
  - COMPRESSION OF GAS PRIOR TO AGR IS ATTRACTIVE. EXTRA COST OF COMPRESSION IS MORE THAN OFFSET BY REDUCTIONS IN COST OF PHYSICAL SOLVENT.
  - FOR SULFUR REMOVAL FROM HIGH-SULFUR COAL GAS, SELECTIVE AGR IS PREFERRED.
- A VARIETY OF APPLICABLE REFERENCE DESIGNS ARE AVAILABLE FOR SELEXOL. THESE CAN BE MODIFIED TO THE NEEDS OF THIS PROJECT.
- PREVIOUS COMPARATIVE EVALUATIONS BY OTHERS TYPICALLY HAVE NOT REVEALED GREAT DIFFERENCES IN THE ANNUAL COSTS OF SELEXOL, RECTISOL, AND BENFIELD WHEN TREATING TO FUEL GAS SPECIFICATIONS.

DECISION: USE SELEXOL WITHOUT COS HYDROLYSIS, BUT WITH CHILLED SOLVENT, FOR 200 PPMV TOTAL SULFUR IN MBG PRODUCT.

Comparative evaluations by others have not revealed any substantial annual cost differences between Selexol, Rectisol, and Benfield when treating to meet fuel gas specifications. Based on this and the above considerations Selexol without COS hydrolysis is selected for System 4. It is noted that meeting a more stringent sulfur removal requirement would require COS hydrolysis or a different process selection.

F. Sulfur Recovery Trades

Sulfur recovery from the H_2S and COS removed in the AGR system is accomplished in System 5, Sulfur Recovery, Alternate Processes considered for this system include Claus plus Scot processes, Claus plus Beavon processes, and Claus plus Wellman-Lord processes. Comparisons considered in selecting a sulfur recovery system are as follows:

ALTERNATE NO.	1	2	3
DESCRIPTION	CLAUS + SCOT	CLAUS + BEAVON	CLAUS + WELLMAN-LORD
MEETS PERFORMANCE CRITERIA	Yes	Yes	Yes
RELATIVE CAPITAL COST	Higher	Lower	Lower
RELATIVE ENERGY CONSUMPTION	Higher	Lower	Lower
PROPRIETARY	Yes	Yes	Yes
COMMERCIALLY PROVEN	Yes	Yes	Yes
APPLIED DEVELOPMENT NEEDS	Slight	Slight	Slight
COULD DELAY IMPLEMENTATION	Yes	Yes	Yes
DATA AVAILABILITY			
SYSTEM-LEVEL DESIGN	Yes	Yes	Yes
COST	Yes	Yes	Yes
SUBSYSTEM LEVEL DESIGN	No	No	No
COST	No	No	No
RELATIVE COMPLEXITY	Higher	Lower	Higher
RELATIVE OPERATING COST	Higher	Lower	Lower
POTENTIAL ENVIRONMENTAL PROBLEM	Some	-	-
BYPRODUCT MARKETABILITY	Yes	Yes	Yes

Evaluation of these considerations led to the following:

- The three processes are generally similar in cost and operating requirements.
- The Scot process becomes more costly than the others when the CO_2/H_2S gas ratio is as high as in coal gasification.
- Scot and Wellman-Lord are slightly more complex than Beavon because they involve a recycle to the Claus plant.

Consideration of the above plus the availability of design information to the design team led to the selection of the Claus plus Beavon system.

G. Steam Generation Trades

Steam is both generated and required by the coal gasification process. In addition, steam potentially may be used as a source of power for prime mover requirements. In addition to process derived steam, an option exists to provide steam with fired boilers. Alternatives considered for this design are no steam boilers, coal fired boilers with flue gas desulfurization, and MBG fired boilers. Comparisons considered are as follows.

STEAM GENERATION COMPARISON

ALTERNATIVE NO.	1	2	3
DESCRIPTION	NO BOILERS	COAL-FIRED WITH FGD	MBG BOILERS
MEETS PERFORMANCE CRITERIA	Yes	Yes	Yes
RELATIVE CAPITAL COST	Base	Much higher	Much higher
RELATIVE ENERGY CONSUMPTION	Base	Higher	Slightly higher
PROPRIETARY	No	Yes*	No
COMMERCIALLY PROVEN	Yes	Yes	Yes
APPLIED DEVELOPMENT NEEDS		NONE	
COULD DELAY IMPLEMENTATION		NO	
DATA AVAILABILITY			
SYSTEM-LEVEL DESIGN	Yes	Yes	Yes
COST	Yes	Yes	Yes
SUBSYSTEM LEVEL DESIGN	Yes	No	Yes
COST	Yes	No	Yes
RELATIVE COMPLEXITY	Base	Much higher	Higher
RELATIVE OPERATING COST	Base	Higher	Much higher
POTENTIAL ENVIRONMENTAL PROBLEM	No	Yes	No
BYPRODUCT MARKETABILITY		QUESTIONABLE	

* FGD is proprietary

Based on the results of the above and the utility trades studies it is concluded that boilers should not be used to generate steam during normal operations. Process derived steam over and above that required for gasification and other process needs should be used to supply prime mover power. Supplemental power is obtained by purchased electricity.

H. Superheater Trades

Having decided to use process derived steam, which is generally saturated, for prime movers it is necessary to consider the use of superheat. The alternatives considered include no superheat, coal fired superheaters with flue gas desulfurization, and MBG fired superheaters. The comparisons considered are as follows:

ALTERNATIVE NO.	1	2	3
DESCRIPTION	DO NOT S.H. PROCESS STEAM	COAL-FIRED WITH FGD	MBG-FIRED
MEETS PERFORMANCE CRITERIA	Yes	Yes	Yes
RELATIVE CAPITAL COST	Lower	Higher	Base
RELATIVE ENERGY CONSUMPTION	Lower	Higher	Base
PROPRIETARY	No	Yes*	No
COMMERCIALLY PROVEN	N/A	No	Yes
APPLIED DEVELOPMENT NEEDS	None	Yes	None
COULD DELAY IMPLEMENTATION	No	Yes	No
DATA AVAILABILITY			
SYSTEM-LEVEL DESIGN	Yes	No	Yes
COST	Yes	No	Yes
SUBSYSTEM LEVEL DESIGN	Yes	No	Yes
COST	Yes	No	Yes
RELATIVE COMPLEXITY	Lower	Higher	Base
RELATIVE OPERATING COST	Lower	Base	Higher
POTENTIAL ENVIRONMENTAL PROBLEM	N/A	Yes	No
BYPRODUCT MARKETABILITY	N/A	Yes	No

* FGD is proprietary

Other engineering studies indicate

- Saturated HP steam generated in gas cooling is a byproduct.
  - About 30% of the available energy in this steam is recoverable via steam turbines.
- Energy input to superheat this steam is 100% recoverable.
- Superheating via waste heat recovery is technically feasible but undesirable due to potential unreliability.
- Superheating via a coal fired superheater is technically feasible but again, undesirable due to potential unreliability.
- Superheating via a MBG fired superheater is technically feasible and offers little or no risk.

- The high operating cost of MBG superheating is offset by:
  1) Higher thermal efficiency in the superheater
  2) Elimination of reliability problems as gas cleanup is not required
  3) Elimination of solid waste disposal problems.

Based on the above considerations the decision is to superheat process derived steam by use of MBG fired boilers.

I. Air Separation Trades

Gasification of coal with the selected technology requires the use of gaseous oxygen. System 6 provides for air separation. Design alternatives include the choice of low pressure gaseous oxygen compression or pumped liquid oxygen and the purity of the oxygen used. Other studies based on 98-99.5 precent oxygen delivered at pressures up to 1200 psig provide the following comparative considerations.

ALTERNATIVE NO.	1	2
DESCRIPTION	LOW PRESSURE GASEOUS OXYGEN COMPRESSION	PUMPED LIQUID OXYGEN
MEETS PERFORMANCE CRITERIA	Yes	Yes
RELATIVE CAPITAL COST	Base	Higher
RELATIVE ENERGY CONSUMPTION	Base	Higher
PROPRIETARY	Yes	Yes
COMMERCIALLY PROVEN	Yes	Yes
APPLIED DEVELOPMENT NEEDS	No	No
COULD DELAY IMPLEMENTATION	No	No
DATA AVAILABILITY		
SYSTEM-LEVEL DESIGN	Yes	Yes
COST	Yes	Yes
SUBSYSTEM LEVEL DESIGN	No	No
COST	No	No
RELATIVE COMPLEXITY	Base	Greater
RELATIVE OPERATING COST	Base	Higher
POTENTIAL ENVIRONMENTAL PROBLEMS	None	None
BYPRODUCT MARKETABILITY	Possible	Possible

In addition to the above considerations a wide variety of prior process design evaluations have indicated that 98 percent low pressure gaseous oxygen compression is preferred. This alternative offers a reduction in operating costs and capital cost without any anticipated penalty in plant performance or reliability.

Therefore based on the above considerations and prior work the air separation system is chosen to produce 98 percent oxygen with compression to gasifier pressure.

J. Water Treatment Trades

Environmental restrictions are such that waste water treatment must be such that effluent discharge to the user must meet stringent requirements. The other alternative is to design the facility for zero discharge.

Composition of the alternatives is as follows:

ALTERNATIVE NO.	1	2
DESCRIPTION	DISCHARGE TO RIVER	ZERO DISCHARGE TO RIVER
MEETS PERFORMANCE CRITERIA	Yes	Yes
RELATIVE CAPITAL COST	Lower	Higher
RELATIVE ENERGY CONSUMPTION	Lower	Higher
PROPRIETARY	No	No
COMMERCIALLY PROVEN	Yes	Yes
APPLIED DEVELOPMENT NEEDS	Possible	Possible
COULD DELAY IMPLEMENTATION	Yes	Yes
DATA AVAILABILITY		
SYSTEM-LEVEL DESIGN	Limited	Yes
COST	Yes	Yes
SUBSYSTEM LEVEL DESIGN	Yes	Yes
COST	Yes	Yes
RELATIVE COMPLEXITY	Lower	Higher
RELATIVE OPERATING COST	Lower	Higher
POTENTIAL ENVIRONMENTAL PROBLEM	Higher	-
BYPRODUCT MARKETABILITY	Lower	Higher
POTENTIAL PROCESS PROBLEMS	Lower	Higher
NON-COMPLIANCE SHUT DOWN POTENTIAL	Yes	No

Evaluation of the above considerations show that

- Zero discharge will result in minimal potential project delays due to environmental interference.
- Need significant data on receiving body flows over a number of years to design discharge system for a plant this size.
- Water treatment sufficient for discharge produces water acceptable for plant use.

Based on these evaluations the zero discharge alternative is chosen.

K. Tar/Oil Disposition Trade

Alternatives considered for the disposition of recovered tars and oils are:

a) burn in fired equipment and
b) sell as a product.

Table III.K.1 presents the considerations relative to each alternative. Based on these considerations, lack of dependable data on product quality/quantity to be expected as a basis for sales and previous team experience with difficulty in finding suitable markets, the decision is to burn the tar/oil product in the boilers of the Steam Generation and Distribution System.

L. Phenol Recovery Trades

The alternatives considered for phenol recovery include:

a) do no recover (bio-oxidation)
b) Phenosolvan process
c) Chempro process.

The considerations applicable to these alternatives are presented in Table III.L.1. Based on these considerations and the fact that:

- Phenol extraction is only needed with Lurgi based facilities and Lurgi generally prefers or insists on use of Phenosolvan
- Direct biological oxidation is very risky due to very high BOD concentration
- Phenosolvan and Chempro are very similar in overall process approach.

The decision is to use Phenosolvan to recover phenols and to burn them in the Steam Generation and Distribution System.

M. Ammonia Recovery Trades

The process alternatives considered for ammonia recovery include:

a) Chevron WWT
b) Phosam-W
c) stripping only

TABLE III.K.1
TAR/OIL DISPOSITION TRADES
LURGI AND BGC/SLAGGER

ALTERNATIVE NO.	1	2
DESCRIPTION	SELL AS BYPRODUCT	USE AS FUEL
MEETS PER ANCE CRITERIA	Yes	Yes
RELATIVE CAPITAL COST	Low	Higher
RELATIVE ENERGY CONSUMPTION	Lowest	Higher
PROPRIETARY	No	No
COMMERCIALLY PROVEN	-	Yes
APPLIED DEVELOPMENT NEEDS	-	-
COULD DELAY IMPLEMENTATION	Yes	No
DATA AVAILABILITY		
SYSTEM-LEVEL DESIGN	Yes	Yes
COST	Yes	Yes
SUBSYSTEM LEVEL DESIGN	Yes	No
COST	Yes	No
RELATIVE COMPLEXITY	Low	Moderate
RELATIVE OPERATING COST	-	-
POTENTIAL ENVIRONMENTAL PROBLEM	-	SOX, NOX
BYPRODUCT MARKETABILITY	Questionable	-

TABLE III.L.1
PHENOL RECOVERY TRADES
LURGI AND BGC/SLAGGER

ALTERNATIVE NO.	1	2	3
DESCRIPTION	DO NOT RECOVER (BIO-OXIDATION)	PHENO-SOLVAN	CHEM-PRO
MEETS PERFORMANCE CRITERIA	Yes	Yes	Yes
RELATIVE CAPITAL COST	Higher	Lower	Lower
RELATIVE ENERGY CONSUMPTION	Higher	Lower	Lower
PROPRIETARY	No	Yes	Yes
COMMERCIALLY PROVEN	Yes	Yes	Yes
APPLIED DEVELOPMENT NEEDS	Some	No	No
COULD DELAY IMPLEMENTATION	Yes	Yes	Yes
DATA AVAILABILITY			
SYSTEM-LEVEL DESIGN	Yes	Yes	Yes
COST	Yes	Yes	Yes
SUBSYSTEM LEVEL DESIGN	Yes	No	No
COST	Yes	No	No
RELATIVE COMPLEXITY	Lower	Higher	Higher
RELATIVE OPERATING COST	Higher	Lower	Lower
POTENTIAL ENVIRONMENTAL PROBLEM	Higher	Lower	Lower
BYPRODUCT MARKETABILITY	No	Possible*	Possible*

* At least will have fuel value

The considerations applicable to these alternatives are presented in Table III.M.1. Based on these considerations and the fact that

- Phosam-W has been found to be the economic choice from past direct comparisons when an ammonia producing gasifier is used
- Power costs are higher for the WWT process
- If ammonia is not recovered severe problems can occur in the sulfur recovery plant
- Phosam-W is proven on dirty gas similar to coal gasification products.

The decision is to use Phosam-W to recover ammonia from sour water.

N. Gas Compression Placement Trades

The K-T gasification process produces a product at near atmospheric pressure. Therefore, compression is required to meet the plant discharge specification of 600 psig. Compression may be done immediately after gas cooling or after acid gas removal or both. The advantages and disadvantages are as follows:

	ADVANTAGE	DISADVANTAGE
- PLACE COMPRESSORS AHEAD OF AGR	LOWER AGR COST	MUST COMPRESS SOUR GAS AND CO_2
- PLACE COMPRESSORS AFTER AGR	COMPRESS SWEET GAS	HIGHER AGR COST
- PLACE SOME COMPRESSION BOTH BEFORE AND AFTER AGR	MEDIUM AGR COST	

Other previous studies by the design team have shown that AGR cost decreases with pressure faster than the cost of compressing H_2S, COS, and CO_2 increases. Therefore, compressing before AGR, after gas cooling, is preferred.

TABLE III.M.1
AMMONIA RECOVERY TRADES

ALTERNATIVE NO.	1	2	3
DESCRIPTION	CHEVRON WWT	PHOSAM -W	STRIPPING ONLY
MEETS PERFORMANCE CRITERIA	Yes	Yes	Yes
RELATIVE CAPITAL COST	Highest	Mid	Lowest
RELATIVE ENERGY CONSUMPTION	Highest	Mid	Lowest
PROPRIETARY	Yes	Yes	No
COMMERCIALLY PROVEN	Yes	Yes	Yes
APPLIED DEVELOPMENT NEEDS	Slight	Slight	Slight
COULD DELAY IMPLEMENTATION	No	No	No
DATA AVAILABILITY			
SYSTEM-LEVEL DESIGN	Yes	Yes	Yes
COST	Yes	Yes	Yes
SUBSYSTEM LEVEL DESIGN	No	No	Yes
COST	No	No	Yes
RELATIVE COMPLEXITY	Highest	Mid	Lowest
ANNUALIZED COST	-	Lowest	-
POTENTIAL ENVIRONMENTAL PROBLEM	None	None	High
BYPRODUCT MARKETABILITY	Yes	Yes	No

O. H_2/CO Ratio-Gas Transportation Study

1. Introduction

In addition to the use of the coal gasification facility product as medium Btu gas, there are potential uses as chemical feedstocks. In almost all cases, chemical feedstock synthesis gas will require adjustment of the H_2/CO ratio produced in the gasification plant by the water gas shift reaction as a first process step. Therefore, the concept of shifting the product gas at the gasification facility prior to transmission requires some consideration as to cost effectiveness and customer accommodation. The concept may be especially interesting to small volume users which would find the economics of scale working against them with regard to shifting within their own battery limits. This study investigates the effect of adjusting the H_2/CO ratio upon the cost of product transportation.

Other studies have investigated hydrogen gas transmission. These studies generally are concerned with transportation of a fixed volume long distances or cross country cost optimization and are not specifically applicable to the case of a plant supplying a product at a given set of conditions and delivering it to more than one customer at a different set of conditions.

The cases investigated in this study include delivery of the gas at 100 miles at pressures of 215 psia and 500 psia; delivery of 50 percent at 50 miles and 50 percent at one hundred miles; delivery of only two modules production from a four module plant; delivery without additional pipeline compression; delivery with one compression station at the fifty mile point; and the effect of initial pressures of 300,500 and 615 psia. Hydrogen to carbon monoxide ratios are varied from that of the base Koppers-Totzek plant product to 100 percent hydrogen. The quantity of gas is not held constant but is varied in accordance with estimates of product produced from a 20,000 TPD plant as a function of the hydrogen to carbon monoxide ratio. Base studies for MBG and alternate products reported in BDM reports BDM/H-80-583-TR and BDM/H-80-481-TR were used as references in determining the volume of gas produced from 20,000 TPD of coal as a function of CO/H_2 ratio.

2. Methodology

The Darcy equation

$$\Delta P/_{100} = \frac{.000336\ f\ W^2}{\rho D^5}$$

with

$\Delta P/_{100}$ = Pressure drop per 100 miles psia

f = friction factor

W = mass of gas lbs

ρ = density - lbs/ft^3

D = diameter - inches

is used throughout this study. This equation was chosen for its simplicity of use in comparative trade study. The compressibility of the product was accounted for by calculation of ρ at an effective pressure. The friction factor is taken from the regular or Moody chart published in numerous places. The effective pressure (P_e) is calculated for pipe sizing purposes to be given by

$$P_e = \frac{2}{3}\left(P_1 + P_2 - \frac{P_1 P_2}{P_1 + P_2}\right)$$

P_1 and P_2 are the initial and final gas pressures.

Brake horsepower is given by

$$BHP = \frac{144}{33000}\left(\frac{K}{K-1}\right) PV \left[R^{\left(\frac{K-1}{K}\right)} - 1\right]$$

where

R = compression ratio

K = ratio of specific heats, C_p/C_v

No allowance has been made for a load factor in determining BHP requirements. However an overall energy efficiency of 85 percent is used for determining electric power costs.

Capital costs have been determined by

$$C_p = C_1 (A W_p + B D + C_r)L$$

$$C_c = C_2 (E H_p + F D)$$

where:

C_1, C_2 = INFLATION FACTORS = 3

C_p = Cost of pipe line

A = Cost of pipe - \$200/ton

W_p = Wt of pipe - Tons/Mile

B = Pipe installation - \$1300/in Mile

D = Pipe diameter - inches

C_r = Right of way costs - \$6000/mi

C_c = Cost of Compressor Station

E = Specific compression costs - \$200/HP

HP = Horsepower = BHP/eff = BHP/.85

F = Compressor Station Costs = \$15000/in diameter

The base numbers shown are for 1965[1]. The inflation factor of 3 is based on the general construction inflation rate for the period 1965-1980.

Total annual costs were calculated by

$$C_o = \frac{0.9 \text{ HP} \times 0.7457 \text{ KW/HP} \times 8760}{0.85} \; 0.01747$$

$$+ .138 (C_p + C_c) + .01 C_p + .03 C_c$$

$$+ 100{,}000$$

The weight of pipe required was based on a tensile strength of 25,000 psi. This is probably a conservative assumption; no other safety factor was used.

1. Khan, A. R. and Panos, P. S., The Economics of Natural Gas Production, Transportation, Storage, and Distribution, Institute of Technology, Chicago, 1965.

3. Cases Studied

The cases studied are described in Table III.0.1.

The quantity of gas transported varied from 1.56M pounds per hour of essentially 100 percent hydrogen to 2,073M pounds per hour of medium Btu gas. Corresponding energy values range from 9,496 MM Btu per hour to 11,426 MM Btu per hour.

Table III.0.1
Case Description

CASE NUMBER	CASE DESCRIPTION
1	Total product gas from the 20,000 TPD K-T plant are delivered at a point 100 miles from the plant at 215 psia. Initial pressure is varied from 300 psia to 615 psia. Product volumes varies in accordance with the H_2/CO ratio which is varied from near 100 percent hydrogen to that produced in the base K-T plant.
2	Total product gas (same as Case 1) at initial pressure equal to 615 psia is delivered at a point fifty miles from the plant at 215 psia. Fifty percent of the gas is transported an additional fifty miles for delivery at 215 psia. A compressor station is included at the fifty mile position.
3	Total product gas (same as Case 1) at an initial pressure of 615 psia is delivered at a point 100 miles away at 500 psia with intermittent compression at the 50 mile position.

4. STUDY RESULTS

Figures III.0.1-3 give the results of Case I as described in Table III.0.1. It is noted that for convenience of graphical presentation that the CO/H_2 ratio is used rather than the inverse H_2/CO ratio. The results show that considerable savings may be realized in transportation shifting the product to higher hydrogen content. On a total plant output basis, hydrogen produced at 300 psia could be transported roughly at the same cost of transporting MBG produced at 615 psia.

Case II examines the effect of delivering low pressure products at an intermediate point along the pipeline. Figures III.0.4-6 show the results of this study. Comparison with Case I indicates that transmission of product gas at low pressure of even a fraction of the product for the last half of the distance should be avoided. Figure six confirms the relatively low cost of transporting hydrogen due to its low mass to energy content ratio. The study also confirms that transporting hydrogen through successive compressor stations should be done at low compression ratios relative to equivalent heating value but heavier gases.

Case III looks at transporting gas at high pressure for delivery 100 miles from the plant at high pressure. This case includes compression at the half way point in line with the results of current gas transmission practices which have shown that optimim compressor stations soacing is 50 to 75 miles. Again the results is that shifting to hydrogen reduces transmission costs.

On a cost per million BTU basis cost of transportation varied from around $0.05 to around $0.11. Thus while the savings obtained by shifting gas composition are substantial on a relative basis, they would not compensate for the higher gas production costs. Thus a large portion of the gas should be destined for chemical rather than fuel use if the effects of composition shifting are to be of net benefit.

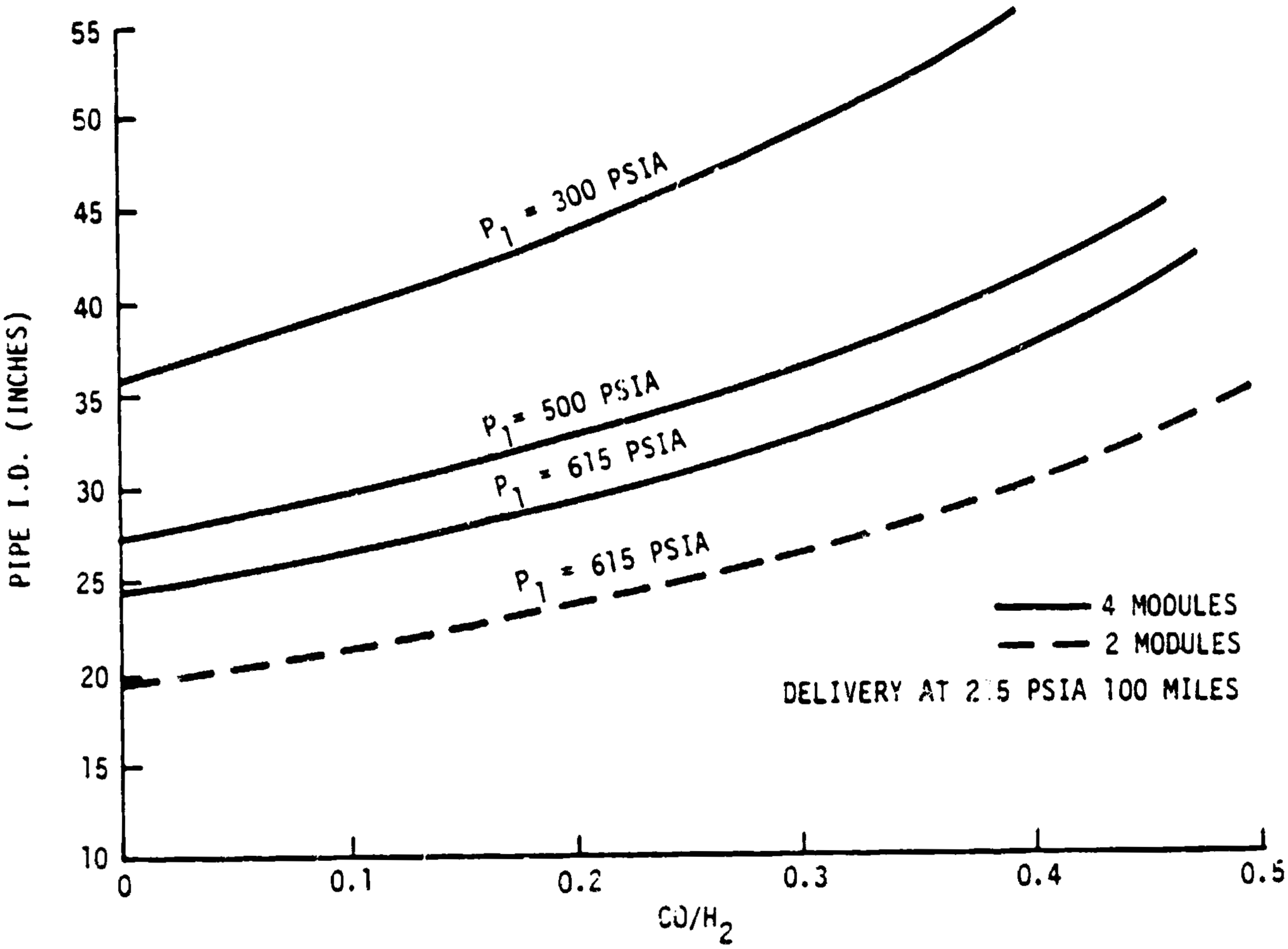


Figure III.O.1. Pipe Diameter Versus CO/H_2 Ratio - Case I

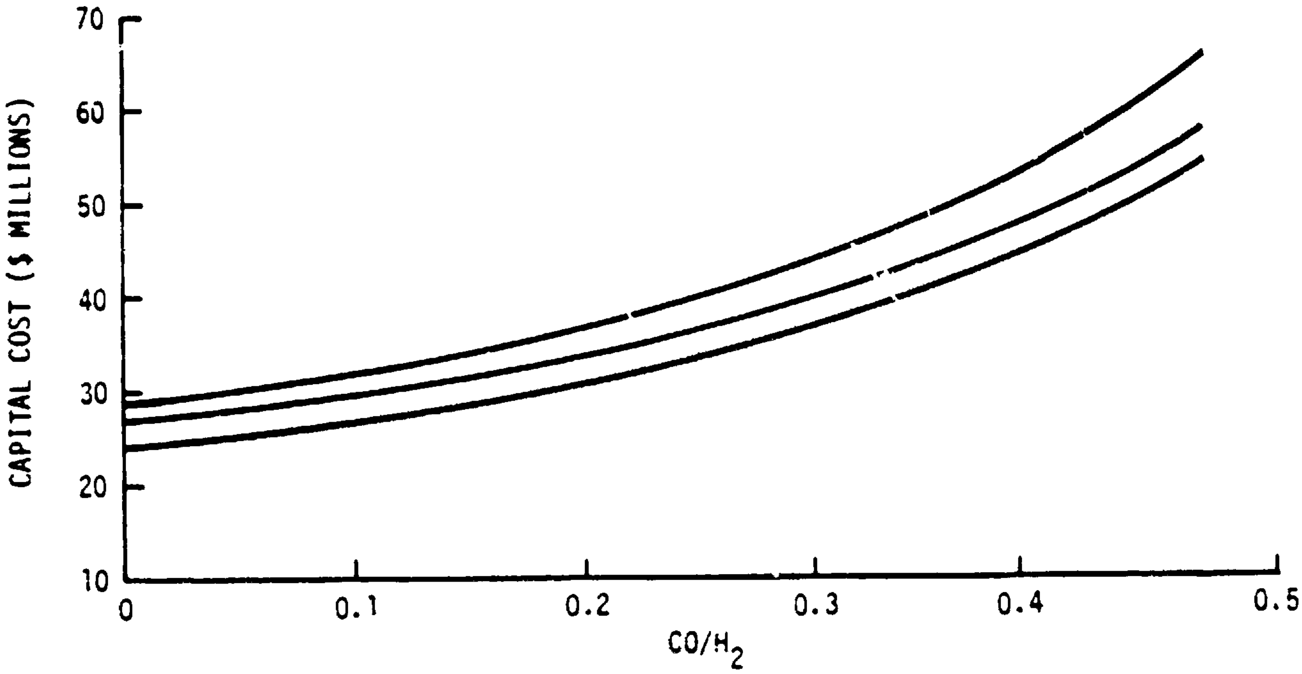


Figure III.0.2. Capital Cost Versus CO/H_2 Ratio - Case 1

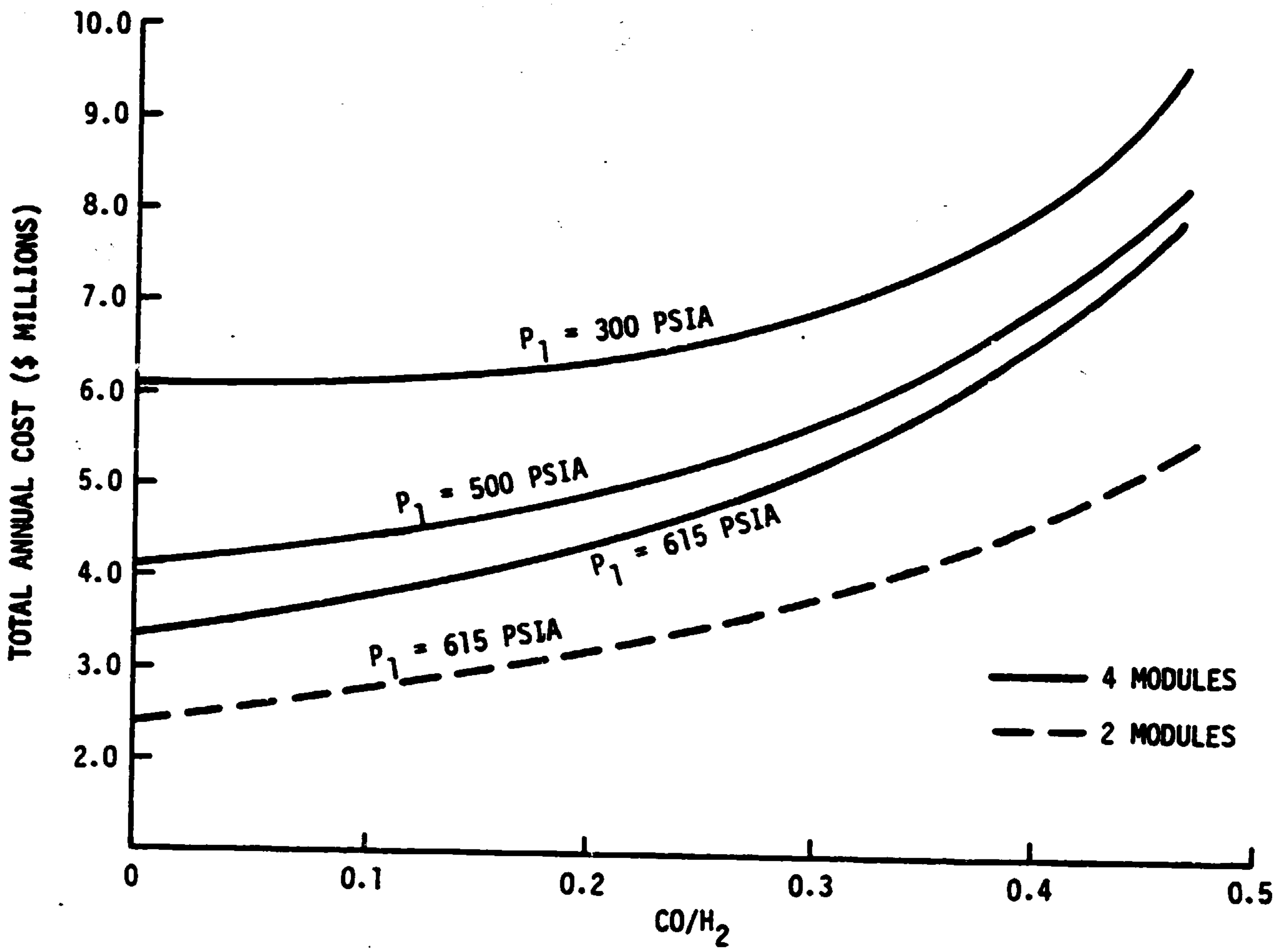


Figure III.0.3. Annual Cost of Service Versus CO/H_2 Ratio - Case I

C-2

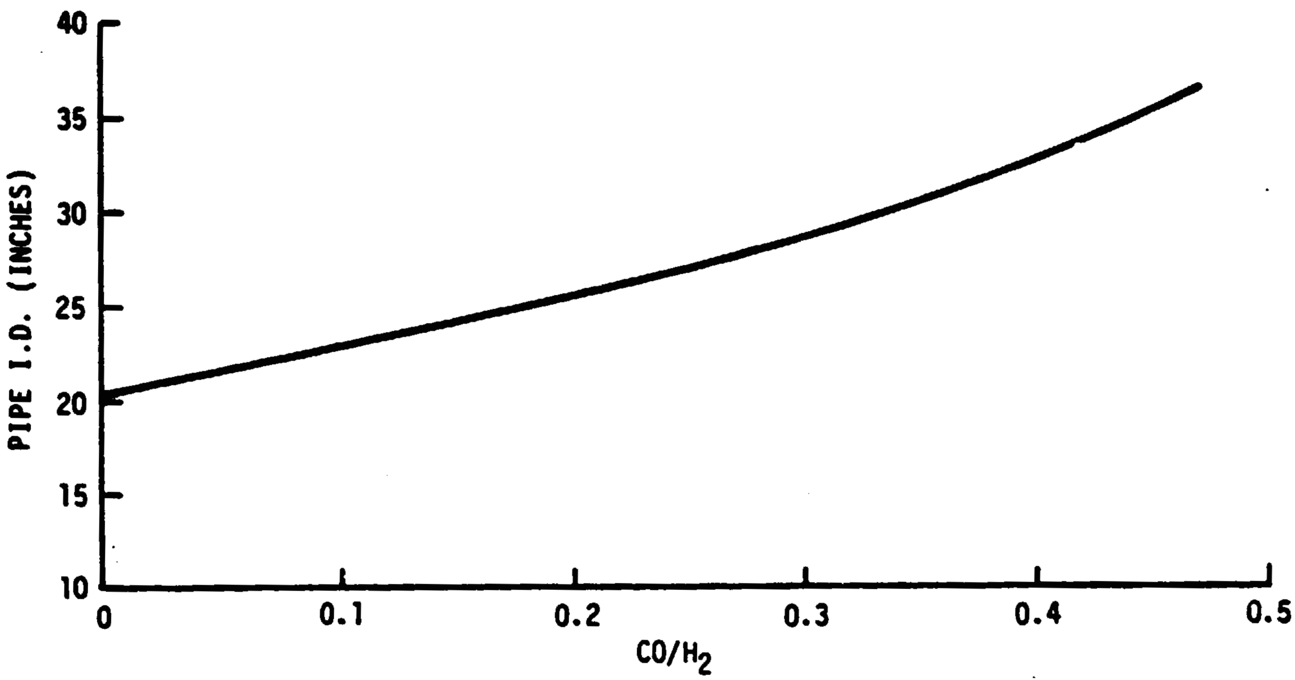


Figure III.0.4. First Leg Pipe Size Requirements - Case II

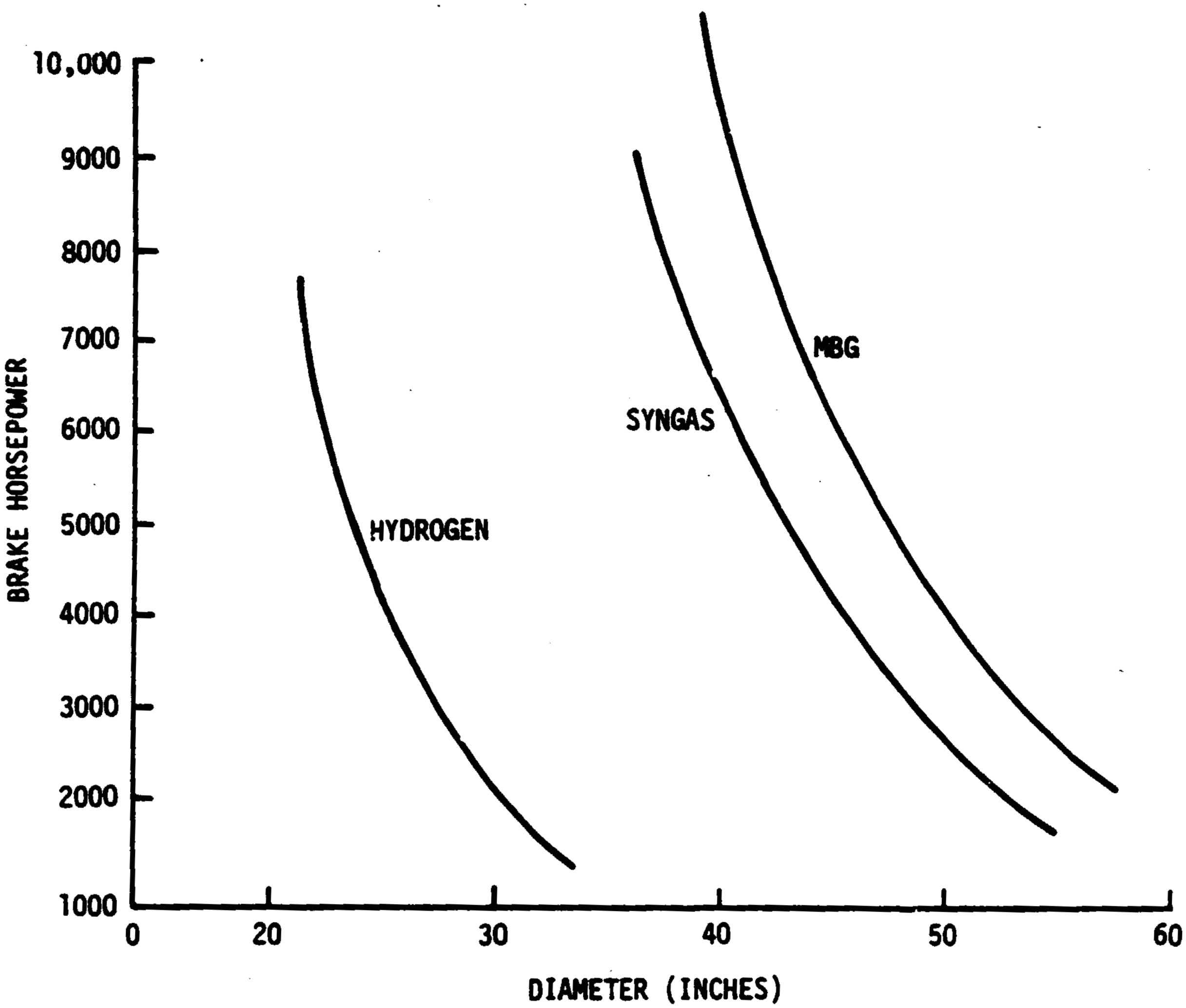


Figure III.0.5. Second Leg Brakehorsepower Requirements - Case II

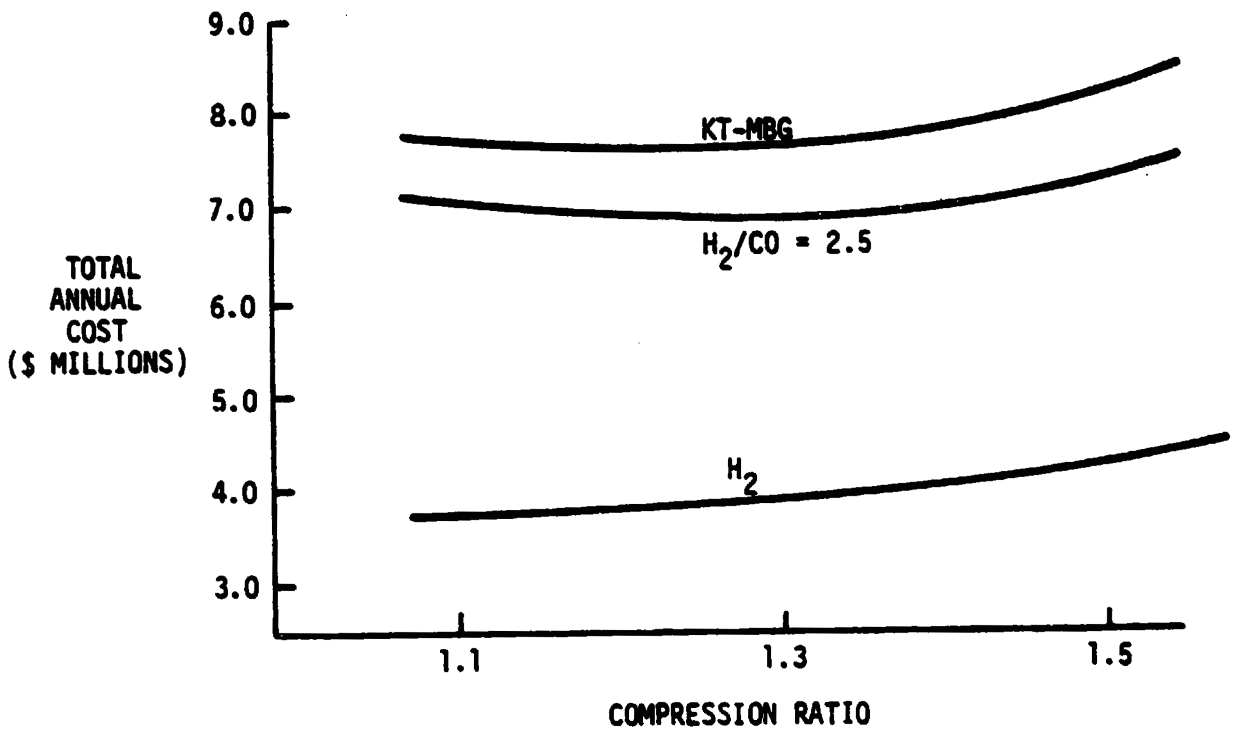


Figure III.0.6. Total Annual Cost of Service Versus Compression Ratio For Second Leg - Case II

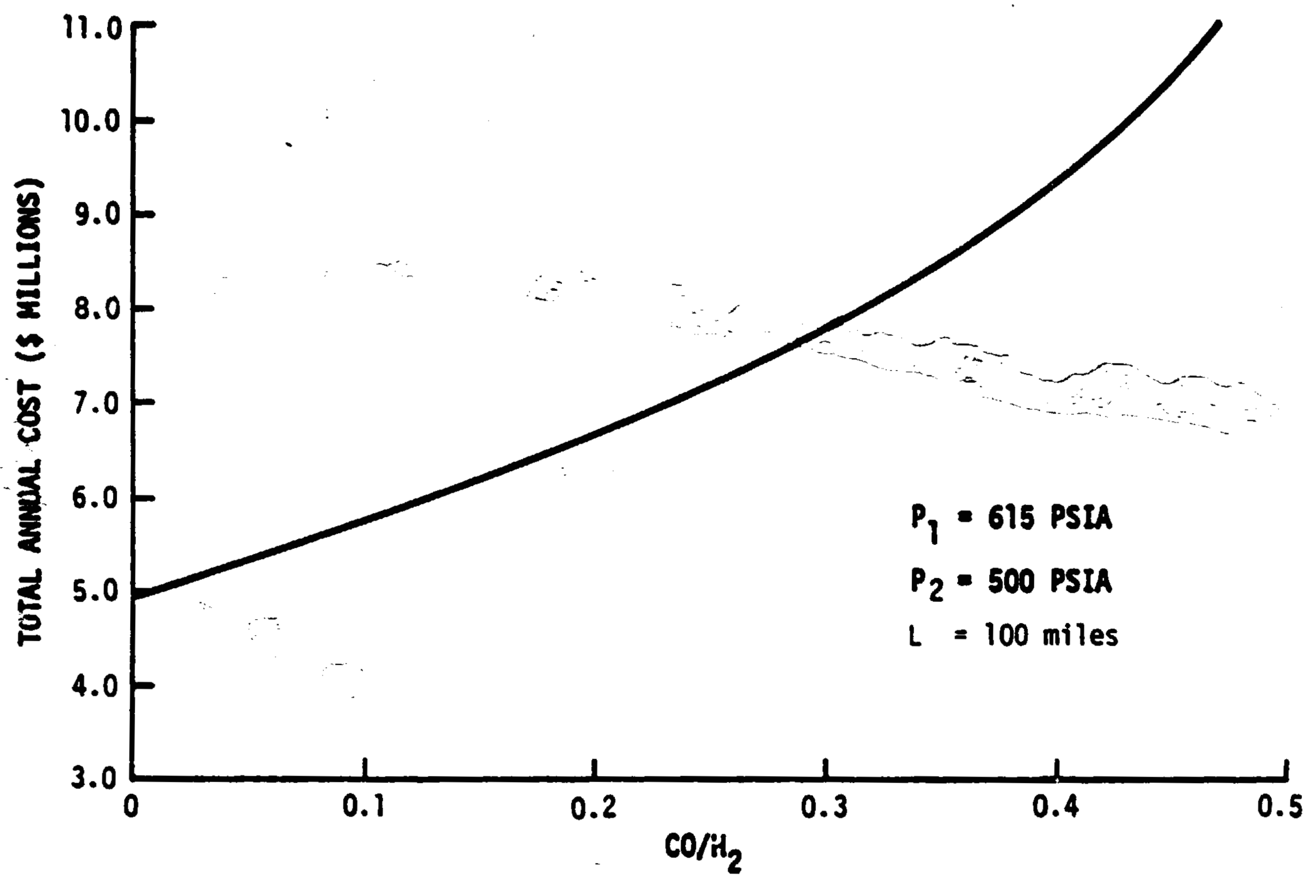


Figure III.0.7. Annual Cost of Service Versus Gas Composition - Case III

CHAPTER IV
MBG FACILITY DESIGN

A. OBJECT

The object of the design task was the development of conceptual and preliminary designs for producing medium Btu gas based on each of the following five coal gasification technologies:

(1) Koppers-Totzek
(2) Texaco
(3) Babcock and Wilcox
(4) Lurgi
(5) BGC/Lurgi

These designs were developed so as to provide a basis for system and unit operation selection, material and energy balances, cost analysis, and schedule analysis.

B. BASIS

Each design presented in this report is based on gasifying 20,000 tons per day of coal in four 5000 TPD modules. Each of the four modules is self-contained and capable of operating alone. Such systems as general facilities, coal receiving and handling, and solids disposal are designed on a total facility basis and serve all four of the operating modules. Other systems, with a few exceptions described later, are sized and dedicated to individual modules. The plants are designed to meet anticipated environmental standards including zero aqueous discharge. The plant products are MBG, sulfur, and, where appropriate, coal fines and ammonia. The design site of the plants is taken to be Murphy Hill, Alabama. Table IV.B.1 lists the outstanding design specifications used in this work as taken from the TVA document, "Design Criteria for Conceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," dated March, 1980.

TABLE IV.B.1. DESIGN BASIS SUMMARY*

PLANT CAPACITY	20,000 TPD
PLANT TYPE	FOUR INDEPENDENT MODULES
COAL TYPE	KENTUCKY NO. 9
PLANT SERVICE FACTOR	90 PER CENT
PLANT LIFE	20 YEARS EACH MODULE
ELECTRICITY SOURCE	TVA GRID-4.16KV, 6.9KV, 13.8KV
WATER	TENNESSEE RIVER
COAL RECEIPT	BARGE +5% BY TRUCK
PRODUCT DELIVERY	MBG AT 600 PSIG
PRODUCT QUALITY	MINIMUM 285 Btu/SCF
SULFUR	PRILLED
COAL COST (1980)	$1.25/MM Btu
LAND COST (1980)	$3,000/ACRE
CLEARING AND GRUBBING (1980)	$2,000/ACRE
ELECTRICITY COSTS (1980)	
BY-PRODUCT CREDIT	NONE
ESCALATION	AS PER TVA SPECIFICATION

*Design Criteria for Cenceptual Designs and Assessments of TVA's Coal Gasification Demonstration Plant," Tennessee Valley Authority, March, 1980

In addition to the primary design coal, provisions were included to allow up to five per cent Alabama coal and up to 2 per cent standard deviations variation in the coal analysis. Tables III.A.1 through Tables III.A.3 in Chapter III contain the analyses of these coals. The primary coal is assumed to be delivered by barge; the Alabama coal is assumed to be delivered by truck. No rail service to the plant is provided.

C. METHODOLOGY

In order to arrive at the most favorable designs possible in a preliminary study, a systematic methodology designed to assure use of all available information was followed.

The open literature was surveyed and all appropriate published design studies were collected for use as a study base. The systems and unit questions required for the designs were arranged in a generalized block flow diagram. Processes and technologies, which are viable candidates for each of the systems, were identified and cataloged (Chapter II and Appendix A). Following the identification of candidate systems and evaluation of the design data from the literature and in-house files, conceptual designs were completed. These designs, called definition level designs, were accomplished by factoring and scaling of published systems. These designs are presented in Appendix B.

With the definition level designs as a basis, a series of trade studies was performed to evaluate the process options available for each system and plant configuration (Chapter III).

Finally, with the plant configurations and preferred processes established, more detailed designs, called reference facility designs, were accomplished according to the design specifications described above.

Reference facility designs were completed based on Koppers-Totzek, Texaco, and Babcock and Wilcox coal gasification technologies. Definition level designs only were developed for the Lurgi and BGC/Lurgi cases. However, the trade studies and reference facility designs for the first three reference facility designs were completed prior to the work with Lurgi and BGC/Lurgi. Therefore, these two latter designs, even though on the conceptual definition level, incorporated the results of the earlier trade studies and more definitive design work where appropriate.

D. GENERAL PROCESS DESCRIPTIONS

Each of the five MBG designs are similar in configuration except where the specific technology dictates otherwise.

Kentucky No. 9 coal is received by barge and either placed in a dead storage pile or processed in System 11, Coal Handling. Provisions have also been made to receive up to five per cent of plant capacity by truck. Dead storage provides for 90 days supply without interruption. Coal, either directly from barges or from dead storage, is processed in a Bradford breaker and conveyed to four process modules of 5000 TPD capacity each.

In each module, the raw coal fed to the plant from System 1 is crushed and dried in System 1 and fed to System 2, gasification section, where it reacts with oxygen from System 6, Air Separation, and steam. Raw gas from Section 2 is processed in System 3, Initial Gas Cleanup and Cooling, where it is cooled to approximately 100^{o}F. Cooled gas is compressed in System 7, if required, and further processed in System 4, Selexol Acid Gas Removal System, where sulfur is reduced to at least 200 PPMV and the heating value is boosted to at least 285 Btu's per cubic foot. Clean gas is dried and delivered at the plant fence for delivery to a pipeline or other chemical processing plant. Table IV.D.1 gives the product from each of the five designs.

TABLE IV.D.1. MBG FACILITY RESULT SUMMARY

PROCESS	KT	TEXACO	B&W	LURGI	BGC/LURGI
NET YIELD (MMSCFD)	900	1,080	976	1,160	959
GAS HHV (BTU/SCF)	305	291	303	308	384
COMPOSITION (VOL. %)					
HYDROGEN	29.6	37.2	30.7	46.8	28.9
NITROGEN	1.5	1.3	3.4	0.4	0.5
CARBON MONOXIDE	63.5	51.2	63.3	17.2	59.6
CARBON DIOXIDE	4.9	9.8	2.6	26.1	1.8
METHANE	0.5	0.5	-	9.0	8.7
ETHANE +	-	-	-	0.4	0.4
(PPM WT.)					
HYDROGEN SULFIDE	62	66	10	101	134
CARBONYL SULFIDE	461	489	58	498	369
WATER	125	102	127	140	134

Systems 10, 11, 12, and 19 serve the total plant and are similar for each of the designs. Other systems are repeated in each module and are described in the next section. Table IV.D.2 lists the current status of these systems.

1. System 10, Instrumentation and Control

The purpose of System 10 is to provide operational monitoring and supervisory master control of module operations. The system includes instrumentation to measure key characteristics of all major streams (mass, temperature, pressure, composition density, and/or others as appropriate) and to display these characteristics in a central location. A dedicated computer is provided to record data of interest at the facility management level. Telecommunications capabilities are provided to communicate with module and system operators as required to control total plant operation to meet delivery requirements. Actual process control is provided at the module or system level as appropriate and is not included in System 10.

2. System 11, Cool Handling

The Coal Handling System provides for the unloading of coal delivered to the plant either by barge or truck, reclaiming the coal from storage, reducing the size of coal, and transporting the coal to Coal Preparation and Feeding, System 1.

Raw coal is unloaded from barges by the barge unloading subsystem which is designed to unload up to ten 1500 ton capacity barges per shift. The coal is unloaded at an average rate of 1200 tons per hour on a 5-day week basis. Coal is transferred by conveyor tc a radial stacker which then produces a kidney-shaped coal pile containing live and dead storage. Coal is reclaimed from live storage and conveyed to a Bradford breaker where it is reduced in size. Coal from trucks is unloaded into a chute from which it is conveyed to the Bradford breaker. Crushed coal from the Bradford breaker is transported to day storage silos from which it is transferred via vibrating belt conveyors to coal preparation and feeding, System 1, for further processing. Coal fines from the Bradford breaker are collected and sent to the gasification unit, System 2.

TABLE IV.D.2. SYSTEM TECHNOLOGY ASSESSMENT

SYSTEM	K-T	TEXACO	B&W	LURGI	BGC-SLAGGER
1 COAL PREPARATION & FEEDING	CP	CA	CA	CP	CP
2 GASIFICATION	CP	RD	RD	CP	RD
3 INITIAL GAS CLEANUP & COOLING	CP	CA/CP	CA/RD	CP	CP
4 ACID GAS REMOVAL	CA	CA	CA	CA	CA
5 SULFUR RECOVERY & TAIL GAS TREATMENT	CA	CA	CA	CA	CA
6 AIR SEPARATION	CP	CP	CP	CP	CP
7 COMPRESSION	CP	-	-	CP	CP
8 PROCESS SOLIDS TREATMENT	CP	CP	CP	CP	CP
9 INCINERATOR	-	-	-	-	-
10 INSTRUMENTATION & CONTROL	CP	CP	CP	CP	CP
11 COAL HANDLING	CP	CP	CP	CP	CP
12 SOLIDS WASTE RECYCLING/DISPOSAL	CP	CP	CP	CP	CP
13 BYPRODUCT PROCESSING	CP	CP	CP	CP/CA	CP/CA
14 PLANT POWER SYSTEM	CP	CP	CP	CP	CP
15 STEAM GENERATION/DISTRIBUTION	CP	CP	CP	CP/CA	CP/CA
16 WATER SUPPLY	CP	CP	CP	CP	CP
17 COOLING WATER SYSTEM	CP	CP	CP	CP	CP
18 WASTE WATER TREATMENT	CA	CA	CA	CA/RD	CA/RD
19 GENERAL FACILITIES	-	-	-	-	-

KEY: CP = COMMERCIALLY PROVEN, CA = COMMERCIALLY AVAILABLE, RD = READY FOR DEMONSTRATION

3. System 12, Solid Waste Recycling/Disposal

The purpose of System 12 is to store solid waste generated by facility operation during the facility life.

This system consists of a lined impounding pit sized to contain 20 years of solid waste. It includes the conveyor system to move the solids to the pit and a leachate recovery area to recover and pump leachate to the process condensate system.

E. KOPPERS-TOTZEK BASED PLANT

A four module, 20,000 TPD, plant based on Koppers-Totzek coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five percent North Alabama coal. Medium Btu gas with heat content of 305 Btu/SCF and not more than 200 ppm sulfur is the primary plant product. Sulfur is recovered for sale as prilled surfur. Ash disposal is on site. The plant is designed for zero water discharge. Trade studies provided the basis for not using boiler produced steam to drive prime movers. Thus, process-derived steam in excess of process requirements is superheated for power use in prime movers. Electricity from the TVA grid is used to supply the balance of the plant prime mover power requirements.

The plant design was arrived at by a systematic procedure based on published design work, process trade studies, team engineering experience, a NASA provided module level definition of twenty systems, and the TVA design criteria document.

The design procedure involved defining available processes to meet the requirements of each system, technical/economic trade studies to select the preferred processes, and engineering design and flow sheet development for each module. Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on stream factor.

The overall plant configuration is shown schematically in Figure IV.E.1. As shown, General Facilities, Instrumentation and Control, Coal Handling, and Solid Disposal serve plant-wide functions. All other plant components are contained in four identical process modules. The instrumentation and control system operates to monitor overall plant performance and to control intermodule relationships. The total list of systems is given in Table IV.E.1.

The results of the design study are given in Table IV.E.2. Appendix B-1 contains the complete design report.

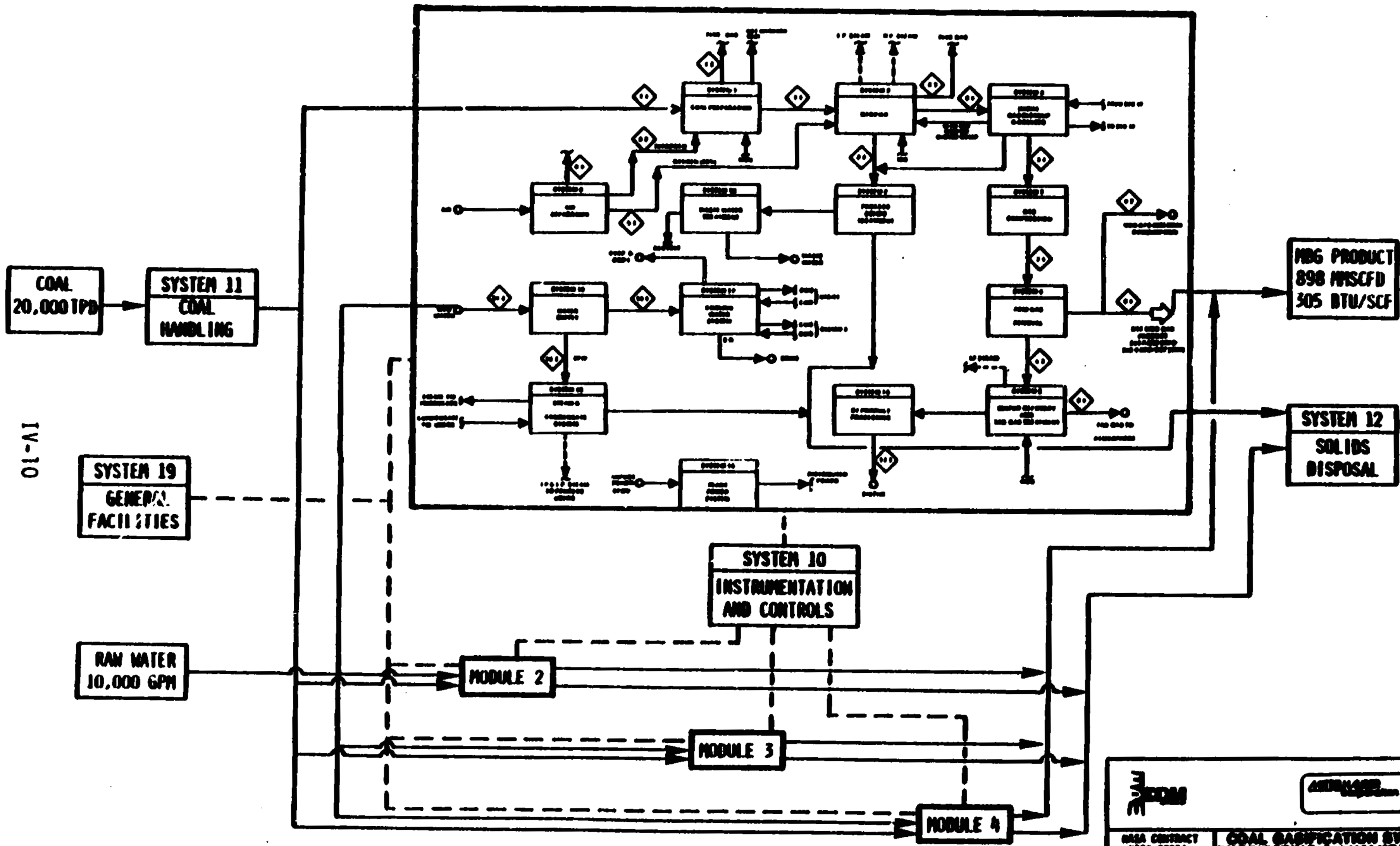


Figure IV.E.1.

COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY

5.2.4

KOPPERS-TOTZEK GASIFICATION REFERENCE FACILITY BLOCK FLOW DIAGRAM

DMD/JVF

824-KT-001

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TABLE IV.E.1. LIST OF SYSTEMS, KOPPERS-TOTZEK PLANT

SYSTEM NO.	NUMBER OF COST UNITS PER MODULE	PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COAL PREPARATION AND FEEDING
2	9	36	GASIFICATION
3	1	4	INITIAL GAS CLEANUP AND COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	2	8	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10	-	1	INSTRUMENTATION AND CONTROL
11	-	1	COAL HANDLING
12	-	1	SOLIDS DISPOSAL
13	1	4	BY-PRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	1	4	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKE-UP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	1	1	GENERAL FACILITIES

TABLE IV.E.2(a). DESIGN STUDY RESULTS*

FEED COAL	6,570,000 TPY
WATER	12,800 GPM
PURCHASED ELECTRICITY	3,720 MM KWHY
MBG PRODUCT	898 M MCFD
MBG PRODUCT	295 MM MSCFY
MBG QUALITY	305 BTU/SCF
SULFUR PRODUCT (PRILLED)	668 LTPD
SULFUR PRODUCT	220,000 LTPY
TOTAL CAPITAL REQUIREMENTS	$2,371 MM
OPERATING AND MAINTENANCE COSTS	$ 47 MM/YR
COAL, CATALYST, CHEMICALS	$ 45 MM/YR
PLANT OPERATING STAFF	346 PERSONS

*Costs are in 1980 Dollars

TABLE IV.E.2(b). CONVERSION EFFICIENCY KOPPERS-TOTZEK PROCESS

INPUTS	10^6 BTU/HR	PERCENT
1. COAL TO FACILITY	18,300	
2. ELECTRIC POWER TO FACILITY	1,448	
OUTPUTS		
3. MBG FROM FACILITY	11,426	
4. COAL FINES FROM FACILITY	-0-	
EFFICIENCY		
COAL-TO-MBG (3 + 1) x 100%		62.4%
OVERALL PRODUCT EFFICIENCY 3 + (1+2) x 100%		57.8%
OVERALL FACILITY EFFICIENCY 3 + 4 (1 + 2) x 100%		57.8%

THE BDM CORPORATION

TABLE IV.E.2(c). OPERATING REQUIREMENTS FOR EXPECTED OPERATIONS
K-T PROCESS - PER MODULE

	BASIS	UNITS
Raw Materials		
Coal	TPY @ 100% Operation	1,825,000 TPY
Catalyst and Chemical Makeup		
	100% Operation	280,200 $/YR
Utility Requirements		
Import Power	kWh/Yr @ 100% Operation	929,896,300 kWh/YR
Operating Requirements		
Labor		
Supervisors	mh/Yr	38,272 Mh/YR
Operators	mh/Yr	130,872 Mh/YR
Supplies	Factored as 15% of Operating Labor Costs	
Maintenance		
Labor	Factored as 1.6% of Total Depreciable Direct Investment	
Supplies	Factored as 2.4% of Total Depreciable Direct Investment	

1. Process Description

The plant consists of eighteen systems including General Facilities. The fourteen systems described here are module level systems and are duplicated four times in the total facility. Table IV.E.3 is an overall material balance for each module.

a. System 1, Coal Preparation and Feeding

This system receives the rough processed coal from coal handling, System 11. The operations carried out in System 1 include: pulverizing coal to the required size, screening and recycle or oversized fractions, drying of processed coal to the required final moisture content, and transporting of the sized, dried coal to Gasification, System 2, for gasification.

Crushed 2" x 0 coal is received from System 11 via a vibrating belt feeder. The coal is weighed on a belt scale after which it goes to a pulverizer-dryer. The pulverizers crush the coal to the required 70% through 200 mesh size, while simultaneously drying the coal to around 2% moisture. Drying is accomplished by a hot nitrogeous gas stream. The pulverizer-dryers are controlled so as to maintain coal particle temperatures between 160-180°F, in order to avoid coking or devolatilization. Transport nitrogen gas is separated from the coal in the service bins followed by dust removal and venting. Gasifier feed bins are connected to the gasifier via variable speed screw feeders. These screw feeders discharge the coal to mixing heads where it is entrained in oxygen and low pressure steam, and transported to the gasifier burners.

b. System 2, Gasification

The Koppers-Totzek gasifier is of the high temperature, co-current entrained flow type. There are four feed points on the four-headed gasifiers, and two burner heads on each point. Each burner is arranged to produce a flame pattern which intersects the pattern of the adjacent burner on a head. Oxygen from the air separation unit, System 6, and low pressure steam generated in the gasifier jacket, transport the pulverized coal to the gasifier.

TABLE IV.E.3. MODULE MATERIAL BALANCE, KOPPERS-TOTZEK PROCESS

(POUNDS PER HOUR)

	RAW COAL	GASIFIER FEED COAL	GASIFIER FEED STEAM	GASIFIER FEED OXYGEN	GASIFIER RAW GAS	COOLED RAW GAS	SWEETENED GAS	NET MBG PRODUCT
CARBON	253,637	253,637	-	-	-	-	-	-
HYDROGEN	17,925	17,925	-	-	16,380	16,380	16,374	14,734
OXYGEN	23,901	23,901	-	336,045	-		-	-
NITROGEN	5,761	5,761	-	6,004	11,764	11,764	11,764	10,587
SULFUR	15,449	15,449	-	-	-	-	-	-
CHLORINE	494	494	-	-	-	-	-	-
ASH	59,650	59,650	-	-	-	-	-	-
CARBON MONOXIDE	-	-	-	-	487,536	487,536	486,660	437,917
CARBON DIOXIDE	-	-	-	-	109,286	109,286	58,718	52,838
METHANE	-	-	-	-	2,005	2,005	2,005	1,804
ETHANE	-	-	-	-	-	-	-	-
LIGHT HC	-	-	-	-	-	-	-	-
TAR + OIL+ NAPTHA	-	-	-	-	-	-	-	-
HCl	-	-	-	-	521	-	-	-
HYDROGEN SULFIDE	-	-	-	-	14,762	14,762	36.0	32.4
CARBONYL SULFIDE	-	-	-	-	2,926	2,926	265.5	238.9
AMMONIA	-	-	-	-	-	-	-	-
HYDROGEN CYANIDE	-	-	-	-	-	-	-	-
TOTAL DRY	376,817	376,817	-	342,049	645,180	644,660	575,822	518,151
WATER	39,853	3,806	56,522	-	186,042	40,435	180	64.8
TOTAL WET	416,670	380,623	56,522	342,049	831,222	685,095	576,002	518,216
T, °F	-	140	270	140	350	100	60	60
P, PSIA	-	-	42.6	21.6	15.4	14.9	630	615

Coal, oxygen, and low pressure steam react to gasify the carbon and volatile matter in the coal. Each module contains eight operating and one spare gasifier train.

Approximately 50 percent of the ash in the coal leaves the gasifier in the form of liquid slag, out the bottom of the gasifier. Entrained slag droplets are quenched with water at the gasifier outlet in order to drop the temperature below the ash fusion temperature. The gas, with entrained particulate ash, passes to the waste heat boiler located directly over the gasifier. The heavier particles of ash are returned via a chute to the slag quench tank below the gasifier. The waste heat boiler cools the gas to about 350^{O}F, prior to clean-up. The waste heat boiler generates 49,500 pounds per hour of 1450 psig steam per gasifier.

The jacket of each gasifier produces 49,600 pounds per hour of low pressure steam from waste heat escaping the gasifier refractory lining. Of this, 7,065 pounds are used as feed to the reactor.

The oxygen feed rate is 0.90 O_2/lb dry coal. The higher heating value of the product gas produced is 305 Btu/SCF.

The Koppers-Totzek gasifiers operate at approximately 7 psig. Each module contains eight operating and one spare reactor.

c. System 3, Initial Gas Clean-Up and Cooling

The Koppers-Totzek gasification system includes a proprietary means for removal of particulate matter, as well as cooling the gas the final amount necessary to achieve temperatures compatible with sulfur removal systems.

The raw gas exiting the waste heat boiler at around 350^{O}F is washed in a pair of Venturi scrubbers arranged in series. The primary scrubber removes the bulk of the particles (around 90%). The gas from the secondary scrubber passes to the final gas cooler. The gas is cooled with water to around 95^{O}F in this cooler. The total particulate removal efficiency of this system is around 99.9%. The cooled gas passes to System 7, compression.

Gas clean-up and cooling are integral parts of the Koppers-Totzek system. Thus, there are eight operating and one spare system in each module.

d. System 7, Compression

The purpose of this unit is to compress the MBG from initial gas clean-up and cooling, System 3, to a pressure sufficient to provide a plant boundary pressure of 600 psig. Water produced in the multi-stage compression process is collected and recycled to System 3 to provide a portion of the make-up required. The product gas compressors for the Koppers-Totzek module are motor driven machines. Outlet pressure from the compression unit is around 650 psig in order to provide for pressure losses in acid gas removal, dehydration, and gas metering.

Each module contains:

(1)	Axial Compressor	34,500 HP
(2)	2-Stage Centrifugal Compressor	34,600 HP
(3)	1-Stage Centrifugal Compressor	21,300 HP

e. System 4, Acid Gas Removal

The acid gas removal system utilizes the Selexol solvent process and receives 645,000 lbs/hr of sour gas at 100^{o}F and 665 psia. The sweet (desulfurized) gas leaves the Selexol absorber at about 615 psia and a temperature of 60^{o}F. The component removals across the absorber are:

	Inlet Gas (mols/hr)	Outlet Gas (mols/hr)	% Removal
Methane	125	125	0.00
Hydrogen	8,125	8,123	0.02
Carbon monoxide	17,406	17,374	0.18
Carbon dioxide	2,483	1,335	46.23
Hydrogen sulfide	433	1	99.77
Carbonyl sulfide	49	4	91.84
Nitrogen	420	420	0.00
Water	42	5	88.10
	28,958	27,387	

The Selexol unit includes an absorption refrigeration unit for chilling the lean solvent so as to achieve the above removals of hydrogen sulfide and carbonyl sulfide. The sour gas from the Selexol unit is rich in CO_2 and at slightly more than atmospheric pressure.

f. System 5, Sulfur Recovery and Tail Gas Treatment

Acid gas from Acid Gas Removal, System 4, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H_2S acid gases. Typically, in a Claus-type sulfur plant, the acid gas is first passed through a knockout drum before entering the reaction furnace. The chemistry of the process involves converting H_2S to elemental sulfur according to the following equation:

$$2H_2S + SO_2 \rightarrow 3S + 2H_2O$$

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in By-Product Processing, System 13, and the tail gas from the final condenser is fed to a Beavon tail gas treating unit where essentially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere. The Beavon sulfur removal process reduces the sulfur content in the tail gas to less than 100 ppm. In this system, hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled and passed into a contactor where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution and transferred to By-Product Processing, System 13. The decanted redox solution is returned to the system.

The system receives 71,597 lbs/hr of acid gas from the Selexol solvent regenerator at about 7 psig and 120°F.

The recovery of by-product elemental sulfur from the sulfur recovery system is 183 short tons/day per module, which amounts to an overall sulfur recovery of about 99.9 percent.

g. System 6, Air Separation

The air separation plant is designed to provide 3,850 short tons/day of 98 percent oxygen for use in the gasification section.

About 1,484,000 lbs/hr of atmospheric air is compressed, in two stages of compression, to a pressure of 110 psia and aftercooled to 100°F. The compressed air is then cryogenically separated into oxygen and nitrogen in a packaged 'cold box'. The separated oxygen from the cold box (at 2 psig and 70°F) is compressed, in four stages of compression, to 40 psia for use in the gasifiers.

Each module contains a 23,000 HP axial air compressor powered by a 1450 psig superheated steam turbine, a 7500 HP centrifugal air compressor driven by a 25 psig steam turbine and a 1750 HP centrifugal oxygen compressor powered by an electrical motor. All turbine drives have electrical start-up motors.

The air separation plant was designed as two trains, each in operating service and each providing one-half of the oxygen requirement.

h. System 8, Process Solids Treatment

Approximately 74,000 pounds per hour per module of ash and slag from the gasifier and gas cooling system along with biological sludges and solid wastes from process condensate treatment are treated in this unit.

Gravity settlers separate dense solids from the waste streams.

Float thickeners-clarifiers treat slurry from the gravity settlers. Thickener and coagulant aids are added to facilitate solid-liquid separation.

Rotary drum filters filter sludges from the process condensate treating systems and the float thickeners.

Recovered water is sent to the process condensate treating system, and solids are conveyed to final solids disposal.

j. System 13, By-Product Processing

Sulfur is the only plant by-product other than ash/char solids which is disposed of on-site. Molten sulfur is pumped to a prilling tower in a continuously flowing circulation system. In the tower, sulfur is dispensed in droplets through nozzles. Droplets fall counter current to a stream of cooling air and solidify prior to landing in the bottom prill collection section. From the prilling tower, sulfur is conveyed to a storage building from which it is transferred by truck or barge for sale.

k. System 14, Plant Power System

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

(1) Develop the necessary voltage stepdown arrangement for plant requirements

(2) Distribute the necessary power to the plant equipment.

TVA's incoming substation transformers receive power from its prevalent distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirement for motors, heaters, lighting, and other miscellaneous loads.

The Medium Voltage Electrical Distribution Systems is a secondary selection system (double ended supply) with several medium voltage buses. Each medium voltage bus receives power from its respective incoming substation transformer through an incoming breaker and supplies power to the medium voltage distribution system through the feeder breakers.

The Low Voltage Electrical Distribution System typically consists of multiple 480 V double-ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Load center transformers are air cooled, dry type, 150°F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consist of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 volt DC system supplies control power for medium voltage and 480 volt plant switchgear control, protective relaying and annunciation. The system also supplies power for emergency lighting.

l. System 15, Steam Generation/Distribution

All of the plant steam is process generated.

Twenty-five psig steam (396,000 pounds per hour per module) is generated in the gasifier jackets. The gasifier steam requirement is 50,000 pounds per hour per module. Steam in excess of the gasifier requirements is used along with steam from System 5, Sulfur Recovery, to drive gas compressor in System 7, the refrigerant cooling water in System 4 and other plant uses. Waste heat boiler steam, generated at 1450 psig, is used to drive gas compressors and a small amount is let down to 550 psig for use in the sulfur recovery system. One hundred psig steam is generated in the sulfur recovery system and from gas compression turbine exhaust. It is used for reboiler heating in System 4 and to deaerate the steam condensate system.

m. System 16, Water Supply

The raw water treatment unit is designed to provide treated and untreated water for the following facility water systems:

(1) Fire water
(2) Service water
(3) Potable water

(4) Cooling water

(5) Boiler feed water

Raw water is pumped from the river to a fire water-raw water storage tank. The total raw water requirement is 4.6 million gallons per day.

The raw water-fire water storage tank provides surge capacity for water treatment as well as storage capacity for fire water. During an emergency, fire water is pumped from the tank to the fire water heater system. The fire water pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the raw water-fire water storage tank to the softener-clarifier. Lime, alum, and polyelectrolyte from the clarifier bulk chemical storage and feed system are added to the softener-clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one wt percent sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the softener-clarifier flows to the self-backwashing sand filters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the sand filter backwash sump and is recycled to the softener-clarifier. The filtered water flows to the filtered water storage tank and is utilized as cooling tower make-up for the process cooling tower as service water for general plant use, as feed to the demineralizer package, and as feed to the potable water system.

Water intended for potable services is chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 ppm free chlorine in the tank.

Filtered water intended as feed to the demineralizer package is injected with sodium sulfide to remove trace amounts of chlorine which adversely affect the demineralizer resins and after filtered through activated carbon to remove any remaining organic contaminants and dissolve iron.

In the demineralizer, the mineral salts present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong cation and magnesium, which the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

The mixed bed polisher is provided to remove silica to 0.02 ppm and to polish returned turbine condensate for reuse.

The boiler feed water deaerating heaters operate at 30 psig and 250°F. The deaerators reduce the oxygen content of BFW to 0.005 cc/liter.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feed water pumps to protect the condensate systems.

n. <u>System 17, Water Cooling</u>

The purpose of this unit is to provide cooling water to the various process users in the facility.

The cooling tower system includes the tower and fans, side stream filters, circulating water pumps, cold water basin, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basin, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

A portion of the circulating water is passed through side stream filters to reduce loading to suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basis is maintained by continuous make-up of the clean water from the raw water treatment system.

The blowdown stream is passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide and is sent to waste treatment for disposal.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algicides are added periodically to further eliminate algae growth. Sulfuric acid is added to control pH, and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally, a polyphosphate dispersant is added to enhance the action of the inhibitors.

o. System 18, Waste Water Treatment

The purpose of this unit is to collect and treat all plant liquid effluent streams. The plant design is predicated on "zero discharge" and permits recycle and reuse of treated water. Streams treated include the following:

(1) Oily water sewers
(2) Coal pile run-off
(3) Storm water run-off
(4) Demineralizer regenerant wastes and rinse water
(5) Cooling tower blowdown
(6) Sanitary waste water
(7) Gasifier slag quench drains
(8) Separated water from solids treatment
(9) Filtrate from biological treatment.

Process operations include:

(1) Oil Separator - streams containing free and dissolved oil and treated in a gravity separator utilizing an emulsion breaking chemical and heat to separate the oil-water mixture

(2) Sour Water Stripper - water streams with appreciable H_2O and HN_3 residuals are steam-stripped to remove these contaminants

(3) Equalization Basin - liquid streams with extremely high or low pH are mixed in an equalizing basis and treated with sulfuric acid or caustic to change the mixed pH to a value of 6.0 - 8.0

(4) Gravity Settling-Thickener - liquid streams with high suspended or dissolved solids are treated in a gravity settler-thickener and mixed with lime, alum, coagulant aids, and polymers to facilitate separation and thickening

(5) Multiple Effect Evaporation - neutralized wastes and brines are evaporated to recover water and concentrate the solids.

The recovered, treated water is used as make-up to cooling towers or raw water supply. The resultant solids are conveyed to the solids disposal system.

p. System 19, General Facilities

The purpose of this unit is to provide equipment or services to support the gasification facility at the facility level.

This unit is a general facility category and provides the following equipment and services:

(1) Administration building
(2) Laboratories
(3) Change rooms
(4) Warehouses
(5) Maintenance buildings
(6) Operation centers
(7) Security offices
(8) Plant air facility
(9) Fire house
(10) Visitor reception
(11) Plant fencing
(12) Plant lighting
(13) Roads, bridges

(14) Docking facilities
(15) Interconnection pipe ways
(16) Fire protection network
(17) Flare stacks and headers
(18) Plant instrument air compressors
(19) Environmental monitoring
(20) Site preparation.

F. TEXACO BASED PLANT

A four module, 20,000 TPD, based on Texaco coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five per cent North Alabama coal. Coal transportation is by river barge except for the Alabama coal which is trucked to the site. Medium Btu gas with heat content of 291 Btu/scf and not more than 200 ppm sulfur is the primary plant product. Sulfur is recovered for sale as prilled sulfur. Ash disposal is on site. The plant is designed for zero water discharge. Trade studies provided the basis for not using boiler produced steam to drive prime movers. Process derived steam in excess of process and plant requirements is used to generate electricity. Electricity from the TVA grid is used to supply the balance of the plant power requirements.

The plant design was arrived at by a systematic design procedure based on published design work, team engineering experience, a NASA-provided module level definition of some twenty systems, and the TVA document, "Design Criteria for Conceptual Designs and Assessments of TVA'a Coal Gasification Demonstration Plant," March, 1980.

The design procedure involved defining available processes to meet the requirements of each system, technical/economic trade studies to select the preferred processes, and engineering design and flow sheet development for each module. Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on stream factor.

The overall plant configuration is shown schematically in Figure IV.F.1. As shown, General Facilities, Instrumentation and Control, Coal Handling, and Solid Disposal serve plant-wide functions. All other plant components are contained in four identical process modules. The Instrumentation and Control System operates to monitor overall plant performance and to control intermodule relationships. The total list of systems is given in Table IV.F.1.

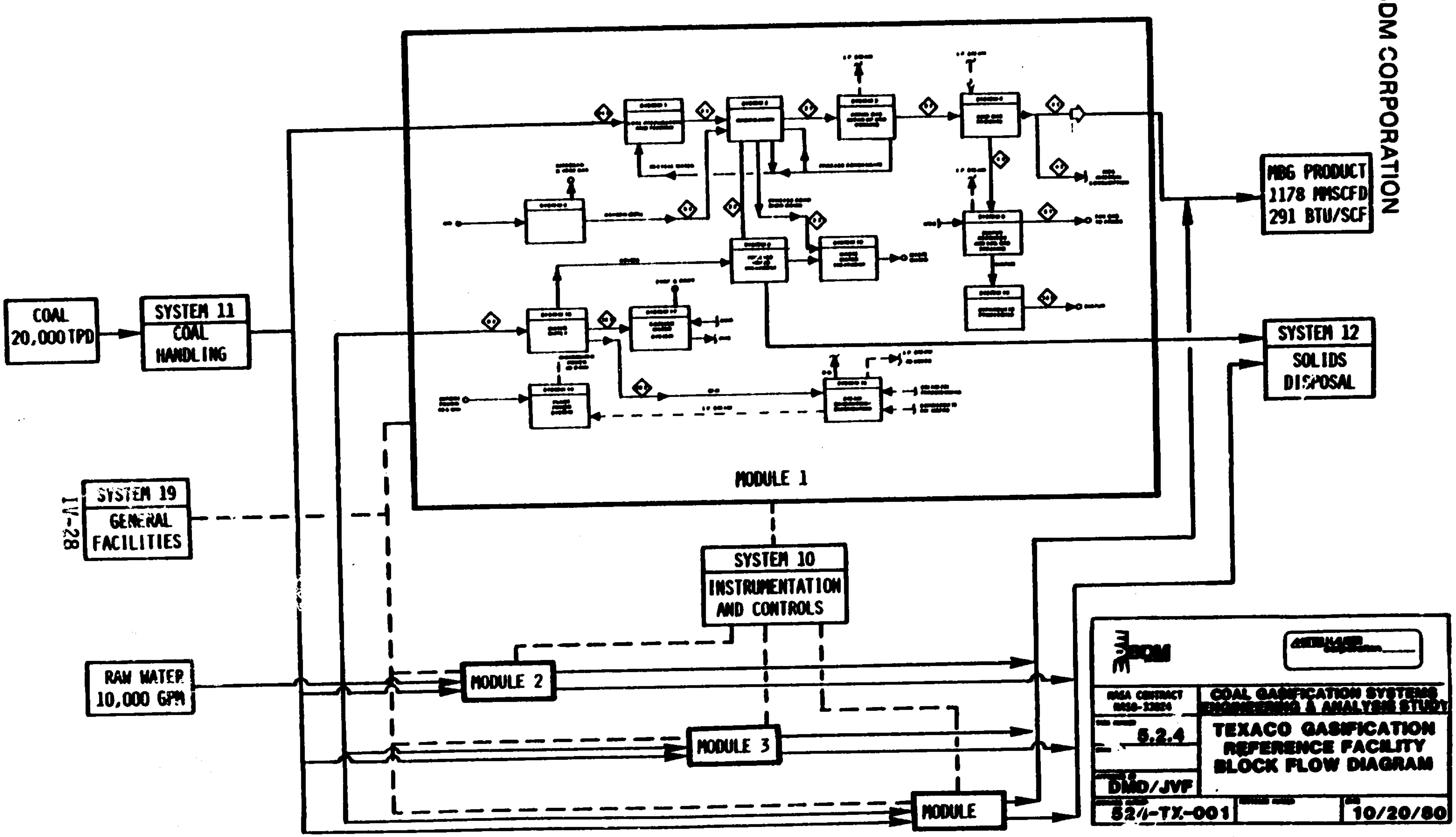

FIGURE IV.F.1. TEXACO PLANT CONFIGURATION

TABLE IV.F.1. MODULE DEFINITION, TEXACO PROCESS

SYSTEM NO.	NUMBER OF COST UNITS PER MODULE	PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COAL PREPARATION AND FEEDING
2	4	16	GASIFICATION
3	1	4	INITIAL GAS CLEANUP AND COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	2 or 3	10	AIR SEPARATION
8	1	4	PROCESS SOLIDS TREATMENT
10	-	1	INSTRUMENTATION AND CONTROL
11	-	1	COAL HANDLING
12	-	1	SOLIDS DISPOSAL
13	1	4	BY-PRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	1	4	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKE-UP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	-	1	GENERAL FACILITIES

The results of the design study are given in Table IV.F.2. Appendix B-2 contains the complete design report.

1. Process Description

The plant consists of eighteen systems including General Facilities. The fourteen systems described here are module level systems and are duplicated four times in the total facility. Exceptions are the air separation system, which has three trains in Module 1 and 3 and two trains in Modules 2 and 4, and the sulfur recovery system, which has two 100 percent trains in Module 1 and one train in Modules 2, 3 and 4.

Raw coal from System 11, Coal Handling, is ground and slurried with water. This slurry, containing 40 percent water, is pumped into three Texaco gasifiers per module operating at 690 psia pressure. The feed is partially combusted with oxygen from the air separation plant heating the mixture to approximately 2450^{o}F. Ash becomes molten slag. Water contained in the feed is vaporized and reacts with the partial combustion products via the steam carbon and water gas shift reactions to produce a product rich in carbon monoxide and hydrogen. Raw gas is quenched within the reactor to 455^{o}F. Molten slag is solidified in granules by the quenching operation and removed by lock hoppering along with quench water. After water-solids separation, the ash is disposed of on-site in a controlled disposal area. In the gas cooling and clean-up system, the gas is further cooled to 100^{o}F before being treated in the acid gas removal system.

Acid gas from System 4, Acid Gas Removal, is processed in System 5, a conventional Claus plant with Beavon Stretford tail gas clean-up. Approximately 184 tons per day of sulfur are produced by each module. Table IV.F.3 is a modular material balance. The clean gas is dried and delivered at 615 psig. Appendix B-2 contains the detailed report. A summary description follows here.

a. System 1, Coal Preparation and Feeding

System 1, Coal Preparation and Feeding, receives 2" x 0 coal feed from System 11, Coal Handling. Each module processes 5,000 TPD of ROM coal in one of two 100 percent units. This system grinds

TABLE IV.F.2(a). DESIGN STUDY RESULTS*

FEED COAL	6,570,000 TPY
WATER	10,132 GPM
PURCHASED ELECTRICITY	1,560 MM KWHY
MBG PRODUCT	1,278 M MCFD
MBG PRODUCT	354 MM MSCFY
MBG QUALITY	291 BTU/SCF
SULFUR PRODUCT (PRILLED)	668 LTPD
SULFUR PRODUCT	220,000 LTPY
TOTAL CAPITAL REQUIREMENTS	$2,091 MM
OPERATING AND MAINTENANCE COSTS	$128 MM/YR
COAL, CATALYST, CHEMICALS	$181 MM/YR
PLANT OPERATING STAFF	271 PERSONS

*Costs are in 1980 Dollars

TABLE IV.F.2(b). CONVERSION EFFICIENCY
TEXACO PROCESS 20,000 TPD FACILITY

INPUTS	10^6 BTU/HR	PERCENT
1. COAL TO FACILITY	18,300	
2. ELECTRIC POWER TO FACILITY	675.6	
OUTPUTS		
3 MBG FROM FACILITY	13,050.8	
4. COAL FINES FROM FACILITY	-0-	
EFFICIENCY		
COAL-TO-MBG (3 ÷ 1) x 100%		71.3%
OVERALL PRODUCT EFFICIENCY 3 ÷ (1 + 2) x 100%		68.8%
OVERALL FACILITY EFFICIENCY 3 + 4 ÷ (1 + 2) x 100%		68.8%

TABLE IV.F.2(c). OPERATING REQUIREMENTS FOR EXPECTED OPERATIONS TEXACO PROCESS - PER MODULE

	BASIS	UNITS
Raw Materials		
Coal	TPY @ 100% Operation	1,825,000 TPY
Raw Water (@ 2533 gpm design)	Gallons/Year @ 100% Operation	1.331×10^9 gallons/year
Catalyst and Chemical Makeup		
Makeup		$255,000/year
Initial		$607,000
Utility Requirements		
Import Power (@ 49500 KW)	KW hr per year @ 100% Operation	433,620,000 KW hr/year
Operating Requirements		
Labor		141,024 mh/yr
Supervisors		29,952 mh/yr
Operators		
Supplies		15% of operating labor
Maintenance Requirements		
Labor		1.5% depreciable direct investment
Supplies		2.4% depreciable direct investment

TABLE IV.F.3. MODULE MATERIAL BALANCE, TEXACO PROCESS

(POUNDS PER HOUR)

	RAW COAL	GASIFIER FEED COAL	GASIFIER FEED STEAM	GASIFIER FEED OXYGEN	GASIFIER RAW GAS	COOLED RAW GAS	SWEETENED GAS	NET MBG PRODUCT
CARBON	253,637	253,637	-	-	-	-	-	-
HYDROGEN	17,925	17,925	-	-	22,252	22,252	22,243	22,108
OXYGEN	23,901	23,901	-	345,212	-	-	-	-
NITROGEN	5,761	5,761	-	6,825	11,153	11,153	11,153	11,085
SULFUR	15,449	15,449	-	-	-	-	-	-
CHLORINE	494	494	-	-	-	-	-	-
ASH	59,650	59,650	-	-	-	-	-	-
CARBON MONOXIDE	-	-	-	-	427,270	427,270	427,172	424,567
CARBON DIOXIDE	-	-	-	-	240,589	240,589	129,279	128,492
METHANE	-	-	-	-	2,474	2,474	2,474	2,459
ETHANE	-	-	-	-	-	-	-	-
LIGHT HC	-	-	-	-	-	-	-	-
TAR + OIL + NAPTHA	-	-	-	-	-	-	-	-
HCl	-	-	-	-	-	-	-	-
HYDROGEN SULFIDE	-	-	-	-	-	15,477	15,477	39
CARBONYL SULFIDE	-	-	-	-	-	1,754	1,754	288
AMMONIA	-	-	-	-	-	1,755	-	-
HYDROGEN CYANIDE	-	-	-	-		-	-	-
TOTAL DRY	376,817	376,817	-	352,037	722,744	720,969	592,648	589,035
WATER	39,853	251,210	-	-	1,198,855	856	198	79
TOTAL WET	416,670	628,027		352,037	1,921,599	721,825	592,846	589,114
T, °F	-	140	-	300	455	100	57	60
P, PSIA	-	-	-	815	690	640	630	615

coal and produces a 60 wt percent coal slurry which is fed to System 2, Coal Gasification. This design is patterned after the one in EPRI report AF-880.

A prescreening system selects the 2" x 1/4" coal for further size reduction in a cage mill. The combined cage mill product and minus 1/4" coal from the prescreening operation are further processed in a wet rod mill. The discharge from the rod mill containing approximately 50 percent water is diluted and screened to produce the desired size consistency with oversized material being recycled to the mill. A centrifuge is used to reduce the water content of the coal slurry. Final water content is controlled at 40 percent by water addition to the centrifuge product in a mix tank. Final slurry control is obtained by density measurement in the mix tank pump around loop. The slurry feed is transferred to and maintained in two parallel agitated tanks prior to transfer to a run tank in System 2, Coal Gasification.

b. System 2, Coal Gasification

The slurry from System 1 is pumped from mix tanks in the grinding and slurry section to the gasifier slurry tank. A circulating pump circulates the slurry through this tank and supplies slurry to the suction of the high pressure charge pumps. The charge pumps raise the slurry pressure sufficient to feed the gasifiers at 690 psia.

The coal-water slurry is fed through a specially developed burner into a refractory-lined gasifier reactor. Partial combustion with oxygen takes place at a pressure of 60 psig, and a temperature of 2450^{o}F to produce a gas consisting mainly of CO, H_2, CO_2, and steam. Most of the sulfur in the coal is converted to H_2S with the balance converted to COS. Nitrogen and argon from the oxygen feed appear in the gas together with most of the nitrogen from the coal. The gas contains a small amount of methane. Some unconverted carbon and all of the ash are removed in the form of slag. The gas is essentially free of uncombined oxygen. Oxygen consumption is 0.92 pounds per pound of dry coal.

The upper section of the gasifier is the refractory-lined chamber in which the partial oxidation reaction takes place.

The gas from the gas generator reaction section passes straight down into the quench section of the gasifier along with the molten ash where it is immediately quenched with water from 2450 to 455°F. Raw gas is sent to System 3, where residual carbon is scrubbed and the gas is further cooled.

Most of the ash in the coal feed agglomerates into essentially carbon-free molten slag droplets, which are quenched and solidified in the lower quench section of the reactor. This slag is settled through the quench water into the lock hopper. The lock hopper is periodically dumped onto a screen, from which the slag is conveyed to the solids treatment system.

Each module consists of three operating and one spare complete trains.

c. System 3, Initial Gas Clean-Up and Cooling

Raw gas at 675 psig and 455°F enter this system from the coal gasification system. The gas is cooled to 380°F in a steam generator producing 110 psig steam. Further cooling to 330°F in a second steam generator produces 50 psig steam. Boiler feed water to the steam generators is heated to 300°F by heat exchange with the gas from the second generator thereby lowering the gas temperature to 297°F. Water condensate is collected in knockout drums following each steam generator and heat exchanger and recycled for quenching in the gasification system. The gas is next processed in a Texaco proprietary scrubbing unit for recovery of soot from the gas stream. The soot-water blowdown from this unit is recycled to System 1, slurry preparation.

Following the gas scrubbing unit, the gas is cooled by air coolers to 140°F. Water condensate is recycled to the gasification quench system and the gas scrubbing unit. The gas is further cooled to 105°F by exchange with cooling water and contacted with the system make-up water in an ammonia scrubber. Water from the ammonia scrubber is used as make-up in the coal slurry preparation unit. Product gas at 100°F and 625 psig is delivered to System 4, Acid Gas Removal.

d. System 7, Compression

The Texaco based plant operates at a sufficiently high pressure to eliminate the requirement of a gas compression system.

e. System 4, Acid Gas Removal System

The acid gas removal system utilizes the Selexol solvent process and receives 772,000 lbs/hr of sour gas at 100°F and 625 psig. The sweet (desulfurized) gas leaves the Selexol absorber at about 620 psig and a maximum temperature of 75½F. The component removals across the absorber are:

	Inlet Gas (mols/hr)	Outlet Gas (mols/hr)	% Removal
Hydrogen	11,037.5	11,033.0	0.04
Carbon monoxide	15,254.2	15,250.7	0.02
Carbon dioxide	5,466.7	2,937.5	46.27
Methane	154.2	154.2	0.00
Hydorgen sulfide	454.2	1.5	99.67
Carbonyl sulfide	29.2	4.8	83.56
Nitrogen	375.0	375.0	0.00
Water	47.5	4.4	90.74
	32,818.5	29,761.1	

The Selexol unit includes an absorption refrigeration unit for chilling the lean solvent so as to achieve the above removals of hydrogen sulfide and carbonyl sulfide. The sour gas from the Selexol unit is rich in CO_2 and at slightly more than atmospheric pressure.

f. System 5, Sulfur Recovery and Tail Gas Treatment

Acid gas from Acid Gas Removal, System 4, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H_2S acid gases. Typically, in a Claus-type sulfur plant, the acid gas is first passed through a knockout drum

before entering the reaction furnace. The chemistry of the process involves converting H_2S to elemental sulfur according to the following equation:

$$2H_2S + SO_2 \quad 3S + 2H_2O$$

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in By-Product Processing, System 13, and the tail gas from the final condenser is fed to a Beavon tail gas treating unit where essentially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere. The Beavon sulfur removal process reduces the sulfur content in the tail gas to less than 100 ppm. In this system, hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled and passed into a contactor where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution and transferred to By-Product Processing, System 13. The decanted redox solution is returned to the system.

The system receives 132,719 lbs/hr of acid gas from the Selexol solvent regenerator at about 7 psig and 120^{o}F.

The recovery of by-product elemental sulfur from the sulfur recovery system is 184 short tons/day per module, which amounts to an overall sulfur recovery of about 99.9 percent.

g. System 6, Air Separation

The air separation plant is designed to provide 4,224 short tons/day per module of 98 percent oxygen for use in the gasification section.

About 1,628,000 lbs/hr of atmospheric air is compressed, in two stages of compression, to a pressure of 110 psia and aftercooled

to 100°F. The compressed air is then cryogenically separated into oxygen and nitrogen in a packaged 'cold box'. The separated oxygen from the cold box (at 2 psig and 70°F) is compressed, in six stages of compression, to 815 psia for use in the gasifiers. Oxygen is left at 300°F for delivery to System 2.

Each module requires an average of 91,000 HP compression. Modules 1 and 3 each have three trains. Modules 2 and 4 each have two trains.

h. System 8, Process Solids Treatment

Approximately 63,000 pounds per hour per module of ash and slag from the gasifier and gas cooling system along with biological sludges and solid wastes from process condensate treatment are treated in this unit.

Gravity settlers separate dense solids from the waste streams.

Float thickeners-clarifiers treat slurry from the gravity settlers. Thickener and coagulant aids are added to facilitate solid-liquid separation.

Rotary drum filters filter sludges from the process condensate treating systems and the float thickeners.

Recovered water is sent to the process condensate treating system, and solids are conveyed to final solids disposal.

i. System 13, By-Product Processing

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k. System 14, Plant Power System

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

(1) Develop the necessary voltage stepdown arrangement for plant requirements

(2) Distribute the necessary power to the plant equipment.

TVA's incoming substation transformers receive power from its prevalent distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirement for motors, heaters, lighting, and other miscellaneous loads.

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The Low Voltage Electrical Distribution System typically consists of multiple 480 V double-ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Load center transformers are air cooled, dry type, 150^{O}F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consist of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 volt DC system supplies control power for medium voltage and 480 volt plant switchgear control, protective relaying and annunciation. The system also supplies power for emergency lighting.

l. System 15, Steam Generation/Distribution

In the absence of a waste heat boiler on the gasifier outlet there is no high pressure steam generated. Gas cooling and the sulfur plant provide 913,000 lb/hr of 125 psia saturated steam used as follows:

(1)	Selexol regenerator	240,000 lb/hr
(2)	Electric power generation	661,000 lb/hr
(3)	Turbine condensate reheat	12,000 lb/hr

Gas cooling and the sulfur plant produce 173,000 lb/hr of 65 psia saturated steam used as follows:

(1)	Selexol unit refrigeration	144,000 lb/hr
(2)	Sulfur heating	2,000 lbs/hr
(3)	Steam tracing	2,000 lbs/hr
(4)	BFW deaeration	25,000 lbs/hr

Steam condensate is recovered at 30 psia and 250°F. Total BFW circulation rate is 1,086,500 lbs/hr.

There are no fired steam boilers contained in the design.

m. System 16, Water Supply

The raw water treatment unit is designed to provide treated and untreated water for the following facility water systems:

(1) Fire water
(2) Service water
(3) Potable water
(4) Cooling water
(5) Boiler feed water

Raw water is pumped from the river to a fire water-raw water storage tank. The total raw water requirement is 3.6 million gallons per day.

The raw water-fire water storage tank provides surge capacity for water treatment as well as storage capacity for fire water. During an emergency, fire water is pumped from the tank to the fire water heater system. The fire water pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the raw water-fire water storage tank to the softener-clarifier. Lime, alum, and polyelectrolyte from the clarifier bulk chemical storage and feed system are added to the softener-clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one wt percent sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the softener-clarifier flows to the self-backwashing sand filters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the sand filter backwash sump and is recycled to the softener-clarifier. The filtered water flows to the filtered water storage tank and is utilized as cooling tower make-up for the process cooling tower as service water for general plant use, as feed to the demineralizer package, and as feed to the potable water system.

Water intended for potable services is chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 ppm free chlorine in the tank.

Filtered water intended as feed to the demineralizer package is injected with sodium sulfide to remove trace amounts of chlorine which adversely affect the demineralizer resins and after filtered through activated carbon to remove any remaining organic contaminants and dissolve iron.

In the demineralizer, the mineral salts present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong cation and magnesium, which the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

The mixed bed polisher is provided to remove silica to 0.02 ppm and to polish returned turbine condensate for reuse.

The boiler feed water deaerating heaters operate at 30 psig and 250^{o}F. The deaerators reduce the oxygen content of BFW to 0.005 cc/liter.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feed water pumps to protect the condensate systems.

n. System 17, Water Cooling

The purpose of this unit is to provide cooling water to the various process users in the facility.

The cooling tower system includes the tower and fans, side stream filters, circulating water pumps, cold water basin, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basin, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

A portion of the circulating water is passed through side stream filters to reduce loading to suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basis is maintained by continuous make-up of the clean water from the raw water treatment system.

The blowdown stream is passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide and is sent to waste treatment for disposal.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algicides are added periodically to further eliminate algae growth. Sulfuric acid is added to control pH, and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally, a polyphosphate dispersant is added to enhance the action of the inhibitors.

o. <u>System 18, Waste Water Treatment</u>

The purpose of this unit is to collect and treat all plant liquid effluent streams. The plant design is predicated on "zero discharge" and permits recycle and reuse of treated water. Streams treated include the following:

(1) Oily water sewers
(2) Coal pile run-off
(3) Storm water run-off
(4) Demineralizer regenerant wastes and rinse water
(5) Cooling tower blowdown
(6) Sanitary waste water
(7) Gasifier slag quench drains
(8) Separated water from solids treatment
(9) Filtrate from biological treatment.

Process operations include:

(1) Oil Separator - streams containing free and dissolved oil and treated in a gravity separator utilizing an emulsion breaking chemical and heat to separate the oil-water mixture

(2) Sour Water Stripper - water streams with appreciable H_2O and HN_3 residuals are steam-stripped to remove these contaminants

(3) Equalization Basin - liquid streams with extremely high or low pH are mixed in an equalizing basis and treated with sulfuric acid or caustic to change the mixed pH to a value of 6.0 - 8.0

(4) Gravity Settling-Thickener - liquid streams with high suspended or dissolved solids are treated in a gravity settler-thickener and mixed with lime, alum, coagulant aids, and polymers to facilitate separation and thickening

(5) Multiple Effect Evaporation - neutralized wastes and brines are evaporated to recover water and concentrate the solids.

The recovered, treated water is used as make-up to cooling towers or raw water supply. The resultant solids are conveyed to the solids disposal system.

p. System 19, General Facilities

The purpose of this unit is to provide equipment or services to support the gasification facility at the facility level.

This unit is a general facility category and provides the following equipment and services:

(1) Administration building
(2) Laboratories
(3) Change rooms
(4) Warehouses
(5) Maintenance buildings
(6) Operation centers
(7) Security offices
(8) Plant air facility
(9) Fire house

(10) Visitor reception
(11) Plant fencing
(12) Plant lighting
(13) Roads, bridges
(14) Docking facilities
(15) Interconnection pipe ways
(16) Fire protection network
(17) Flare stacks and headers
(18) Plant instrument air compressors
(19) Environmental monitoring
(20) Site preparation.

G. BABCOCK AND WILCOX BASED PLANT

A four module, 20,000 TPD, based on B&W coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five per cent North Alabama coal. Coal transportation is by river barge except for the Alabama coal which is trucked to the site. Medium Btu gas with heat content of 303 Btu/scf and not more than 200 ppm sulfur is the primary plant product. Sulfur is recovered for sale as prilled sulfur. Ash disposal is on site. The plant is designed for zero water discharge. Trade studies provided the basis for not using boiler produced steam to drive prime movers. Thus, process-derived steam in excess of process requirements is superheated for power use in prime movers. Electricity from the TVA grid is used to supply the balance of the plant prime mover power requirements.

The plant design was arrived at by a systematic design procedure based on published design work, team engineering experience, a NASA-provided module level definition of some twenty systems, and the TVA document, "Design Criteria for Conceptual Designs and Assessments of TVA'a Coal Gasification Demonstration Plant," March, 1980.

The design procedure involved defining available processes to meet the requirements of each system, technical/economic trade studies to select the preferred processes, and engineering design and flow sheet development for each module. Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on stream factor.

The overall plant configuration is shown schematically in Figure IV.G.1. As shown, General Facilities, Instrumentation and Control, Coal Handling, and Solid Disposal serve plant-wide functions. All other plant components are contained in four identical process modules. The Instrumentation and Control System operates to monitor overall plant performance and to control intermodule relationships. The total list of systems is given in Table IV.G.1. Design results are presented in Table IV.G.2. The complete design report is given in Appendix B-3.

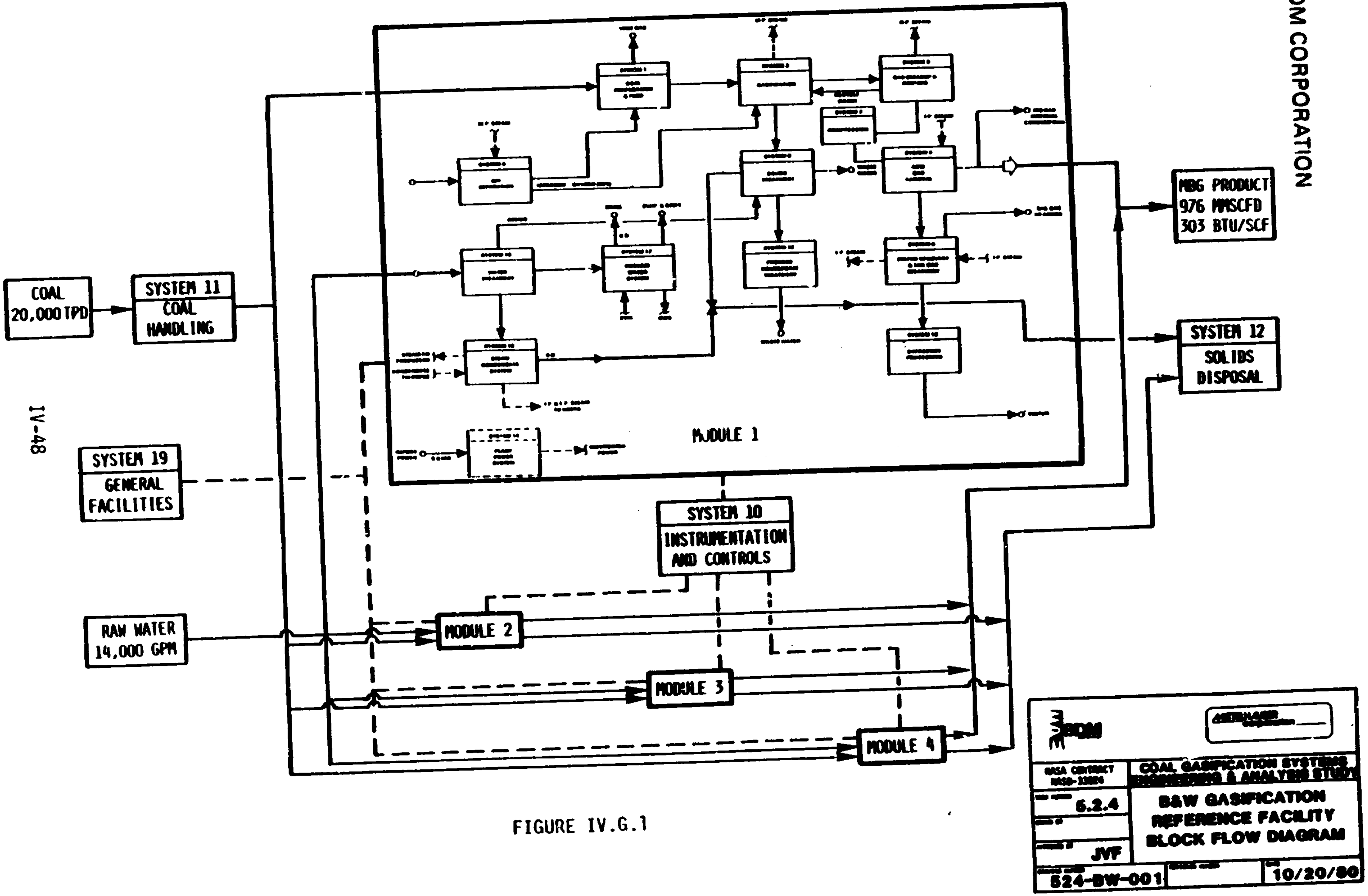
COAL
20,000 TPD
SYSTEM 11
COAL HANDLING
SYSTEM 19
GENERAL FACILITIES
RAW WATER
14,000 GPM
MODULE 1
SYSTEM 10
INSTRUMENTATION AND CONTROLS
MODULE 2
MODULE 3
MODULE 4
MBG PRODUCT
976 MMSCFD
303 BTU/SCF
SYSTEM 12
SOLIDS DISPOSAL
BDM
NASA CONTRACT
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
5.2.4
B&W GASIFICATION REFERENCE FACILITY BLOCK FLOW DIAGRAM
JVF
524-BW-001
10/20/80

FIGURE IV.G.1

TABLE IV.G.1. LIST OF SYSTEMS, B&W PLANT

SYSTEM NO.	NUMBER OF COST UNITS PER MODULE	PER FACILITY	SYSTEM DESCRIPTION
1	3	12	COAL PREPARATION AND FEEDING
2	3	12	GASIFICATION
3	3	12	INITIAL GAS CLEANUP AND COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	2	8	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10	-	1	INSTRUMENTATION AND CONTROL
11	-	1	COAL HANDLING
12	-	4	SOLIDS DISPOSAL
13	1	4	BY-PRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	1	4	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKE-UP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	-	1	GENERAL FACILITIES

TABLE IV.G.2(a). DESIGN STUDY RESULTS*

FEED COAL	6,570,000 TPY
WATER	14,000 GPM
PURCHASED ELECTRICITY	183 MM KWHY
MBG PRODUCT	976 M MSCFD
MBG PRODUCT	322 MM MSCFY
MBG QUALITY	303 BTU/SCF
SULFUR PRODUCT (PRILLED)	673 LTPD
SULFUR PRODUCT	222,000 LTPY
TOTAL CAPITAL REQUIREMENTS**	$3,347 MM ($2,567 MM)
OPERATING AND MAINTENANCE COSTS	$ 138 MM/YR
COAL, CATALYST, CHEMICALS	$ 181 MM/YR
PLANT OPERATING STAFF	271 PERSONS

*COSTS ARE IN 1980 DOLLARS

**BASED ON INSTALLATION FACTOR OF 2.31 AND 1.5. (SEE CHAPTER V.A.)

TABLE IV.G.2(b)
CONVERSION EFFICIENCY
BABCOCK & WILCOX PROCESS
20,000 TPD FACILITY

	10^6 BTU/HR	PERCENT
INPUTS		
1. COAL TO FACILITY	18,300	
2. ELECTRIC POWER TO FACILITY	19.8	
OUTPUTS		
3. MBG FROM FACILITY	12,235	
4. COAL FINES FROM FACILITY	-0-	
EFFICIENCY		
COAL-TO-MBG (③ ÷ ①) x 100%		66.9%
OVERALL PRODUCT EFFICIENCY ③ ÷ (① + ②) x 100%		66.8%
OVERALL FACILITY EFFICIENCY ③ + ④ ÷ (① + ②) x 100%		66.8%

1. Process Description

The plant consists of eighteen systems including General Facilities. The fourteen systems described here are module level systems and are duplicated four times in the total facility. Table IV.G.3 presents the material balance for the plant.

a. System 1, Coal Preparation and Feeding

The coal preparation system receives 5,000 tons/day of crushed, raw coal ranging in size from 1.25" to 0". The raw coal is first pulverized (to 70% passing through a 200 mesh screen) and dried. Drying is accomplished by sweeping hot flue gas, at 450°F, through the coal pulverizer. The dry, pulverized coal is separated from the flue gas in a cyclone recovering about 80 percent of the coal. The remaining 20 percent of the pulverized coal (fines) in the flue gas is recovered by venting the flue gas through a fabric filter baghouse.

Since the gasifier operates at a pressure of 240 psia, it is necessary to pressurize the coal to about 290 psia for transport and injection into the gasifier. Two lock hoppers, each of 20 minutes storage, are pressured with compressed nitrogen. The pressurized coal from the lock hoppers is fed into a pressurized coal feed tank of about 40 minutes storage, from which the pulverized coal is transported to the gasifier using a portion of the compressed nitrogen as the transport medium.

Coal from the gasifier coal feed tank is continuously transported, using compressed nitrogen, into the gasifier.

b. System 2, Gasification

The gasification system includes coal, oxygen and recycle char injection nozzles, the gasifier vessel, slag removal equipment and a steam drum system.

The gasifier vessel consists of a vertical, cylindrical outer steel shell with an inner shell of water-cooled tubes (water wall) in which steam is generated by heat transferred into the tubes from the gasification zone. In the hot reaction zone (lower part of the gasifier),

TABLE IV.G.3. MODULE MATERIAL BALANCE, B&W PROCESS

(POUNDS PER HOUR)

	RAW COAL	GASIFIER FEED COAL	GASIFIER FEED STEAM	GASIFIER FEED OXYGEN	GASIFIER RAW GAS	COOLED RAW GAS	SWEETENED GAS	NET MBG PRODUCT
CARBON	253,637	253,637			51,508			
HYDROGEN	17,925	17,925			18,699	18,699	18,691	16,614
OXYGEN	23,900	23,900		314,914				
NITROGEN	5,761	23,253		8,012	31,252	31,252	31,252	27,779
SULFUR	15,450	15,450						
CHLORINE	492	492						
ASH	59,650	59,650			112,864			
CARBON MONOXIDE					535,414	535,414	515,275	455,779
CARBON DIOXIDE					63,258	63,258	33,967	30,191
METHANE								
ETHANE								
LIGHT HC								
TAR + OIL+ NAPTHA								
HCl								
HYDROGEN SULFIDE					15,385	15,385	7	6
CARBONYL SULFIDE					1,829	1,829	36	32
AMMONIA					16			
HYDROGEN CYANIDE					506			
TOTAL DRY	376,817	394,307		322,926	830,731	665,837	619,228	550,401
WATER	39,850	7,693	32,700		25,190	820	155	70
TOTAL WET	416,667	402,000	32,700	322,926	855,921	666,657	619,383	550,471
T,°F		100		77	1,800	100	60	60
P,PSIA		290	350	240	240	230	615	615

the tubes are covered with a dense refractory to protect the tubes from the molten, flowing slag and yet permit the high temperature required to maintain the gasification reactions and to maintain the slag as a molten fluid. Above the hot reaction zone, the gasification reactions are essentially completed and molten slag is not present. Thus, the inner shell of water-cooled tubes in the upper part of the gasifier is left bare (no refractory covering) to maximize heat transfer from the hot raw gas product into the tubes. The raw gas leaves the gasifier top at 1,800^{0}F and 200 psia, and the molten slag leaves the gasifier bottom at 3,000^{0}F.

The molten slag drains continuously from the gasifier bottom and is quenched and shattered in a water-filled, pressurized slag quench tank. At intermittent intervals, the quenched slag is drained from the quench tank into a slag lock hopper from which the slag is sluiced to dewatering facilities.

The steam generated in the gasifier tubes is made available for use within the plant.

c. <u>System 3, Initial Gas Cleanup and Cooling</u>

Facilities are provided to recover char (coal ash and unconverted coal) from the raw product gas and to recycle it into the gasifier. Two cyclone stages are provided for that purpose. The hot gas from the gasifier flows through the primary char recovery cyclone stage at 1800^{0}F and is cooled down to 450^{0}F in the WHB (waste heat recovery boiler). The 450^{0}F gas then flows through the secondary char recovery cyclone stage. The collected char from the primary and secondary cyclones is transported back into the gasifier injection nozzles by using steam as the transport medium. The 1800^{0}F char from the primary cyclone passes through a char cooler and is cooled to 1200^{0}F. The recycle char mixture of primary cyclone char (1200^{0}F), secondary cyclone char (450^{0}F) and steam has a temperature of about 900^{0}F entering the gasifiers.

The WHB is designed to remove sensible heat from the hot raw gas as it is cooled from 1800°F down to 450°F and convert the recovered heat into steam.

The gas passes through a Venturi scrubber and is cooled to 100°F, which is below its water dewpoint and, therefore, water is condensed out of the gas (i.e., the gas is dehumidified). At the same time, the Venturi scrubber removes essentially all of the residual char particulates from the gas. The clean, cool gas then flows to the gas compressors.

The performance of char removal facilities in the gas cleaning and cooling train may be summarized as:

(1)	Primary Char Cyclone:	Removes 85% of the char in the gasifier outlet gas
(2)	Secondary Char Cyclone:	Removes 33% of the residual char (5% of the original char content)
(3)	Venturi Scrubber:	Removes essentially 100% of the residual char (10% of the original char content).

During the work on this project, preliminary design information was made available by the Babcock & Wilcox Company. That information is referred to herein as the 'B&W design', which compares in feed rate with the design herein as follows:

	This Design	B&W Design
Raw coal feed, T/D	5,000	5,368
Raw coal moisture, %	9.6	6.9
Dry coal feed, T/D	4,824	5,000
Dry coal moisture, %	2.0	0.0

The coal drying equipment was designed to reduce the water content of the coal from 9.6 percent in the raw coal to 2.0 percent in the dry, pulverized coal. Removing that amount of water, and heating the coal to 150°F, requires about 50 million Btu/hr of heat including an allowance for heat loss. The amount of flue gas, entering the pulverizer at 450°F and leaving at 150°F, must be about 695,000 lbs/hr to transfer

50 million Btu/hr of heat. To generate a flue gas at 450°F, the direct-fired air heater must be supplied with about 10.5 times the stoichiometric combustion air needed to burn the MBG fuel gas. Thus, the required air rate to the heater is 142,000 SCFM and the fuel required is 70 million Btu/hr.

A material balance around the gasifier was developed by prorating from the B&W design. Based on that material balance, the carbon conversion in the gasifier was calculated to be 97.5 percent, which coincides with B&W's design information for the selected oxygen-to-coal ratio. The material balance was checked to see that each component (carbon, oxygen, nitrogen, hydrogen and ash) also balanced.

Having confirmed the enthalpy data base, the slag quench duty of 42 million Btu/hr and the recycle char cooling duty of 25 million Btu/hr were calculated by using that enthalpy data base.

d. System 8, Process Solids Treatment

This section describes and discusses the design of the gasification section cooling system, which supplies the integrated cooling requirements for quenching the gasifier slag and for supplying part of the Venturi scrubber cooling. The integrated gasification water loop functions to recover and reuse the slag sluice water as well as to cool the water. The clarifier in the system serves to dewater the process solids by-product before transporting to disposal.

The slag sluice water is heated by absorbing the slag quenching duty. Since the water is recycled for reuse, it is cooled to remove the heat absorbed and the slag solids are removed from the water.

The residual char removed from the raw gas in the Venturi scrubber amounts to 16,381 lbs/hr, which exits with the water leaving the bottom of the scrubber. That water joins the slag sluice water containing 49,675 lbs/hr of the quenched and shattered slag. Thus, the total solids to be removed in the clarifier amounts to 66,066 lbs/hr. Assuming that the sludge from the clarifier is 25 wt percent water, the sludge contains 198,200 lbs/hr of water (the cooling system blowdown) and 66,066 lbs/hr of solids.

e. System 6, Air Separation

The air separation plant is designed to provide 3,850 short tons/day of 98 percent oxygen for use in the gasification section.

About 1,484,000 lbs/hr of atmospheric air is compressed, in two stages of compression, to a pressure of 110 psia and aftercooled to 100°F. The compressed air is then cryogenically separated into oxygen and nitrogen in a packaged 'cold box'. The separated oxygen from the cold box (at 2 psig and 70°F) is compressed, in four stages of compression, to 290 psia for use in the gasifiers.

A part of the separated nitrogen (about 34,600 lbs/hr at 2 psig and 70°F) is compressed, in four stages of compression, to 295 psia for use in pressurizing the gasifier coal feed system and in transporting coal into the gasifiers.

The air separation plant was designed as two trains, each in operating service and each providing 1,925 tons/day of oxygen. However, all of the quantities referred to in this discussion are for the full 3,850 tons/ day of oxygen output per module.

f. System 7, Raw Gas Compression

The raw gas from the Venturi scrubber (666,000 lbs/hr at 100°F and 200 psia) is compressed, in two stages of compression, to 640 psia and aftercooled to 100°F for processing through the acid gas removal system (Selexol unit).

The interstage cooling down to 130°F is provided by air-cooling. Aftercooling is provided by air-cooling down to 120°F and by water-cooling down to 100°F.

The raw gas compression train was designed as a single, operating train with no standby train.

The compression requirements for the 5,000 T/D coal gasification module may be summarized as:

Air compression	72,700 HP
Oxygen compression	19,200 HP
Nitrogen compression	2,430 HP
Raw gas compression	23,000 HP

As shown in the steam balance section herein, all of the compressors are driven by steam turbines. The air, oxygen and raw gas compressors are also provided by standby motors for use during plant start-up before steam becomes fully available.

g. System 4, Acid Gas Removal System

The acid gas removal system utilizes the Selexol solvent process and receives 666,000 lbs/hr of sour gas at 100°F and 635 psia. The sweet (desulfurized) gas leaves the Selexol absorber at about 630 psia and a maximum temperature of 75°F. The component removals across the absorber are:

	Inlet Gas (mols/hr)	Outlet Gas (mols/hr)	% Removal
Hydrogen	9,275.5	9,271.5	0.04
Carbon monoxide	19,115.1	19,110.1	0.03
Carbon dioxide	1,437.1	771.8	46.29
Hydrogen sulfide	451.5	0.2	99.96
Carbonyl sulfide	30.9	0.5	98.38
Nitrogen	1,030.2	1,030.2	0.00
Water	45.5	4.5	90.11
	31,385.8	30,188.8	

The Selexol unit includes an absorption refrigeration unit for chilling the lean solvent so as to achieve the above removals of hydrogen sulfide and carbonyl sulfide. The sour gas from the Selexol unit is rich in CO_2 and at slightly more than atmospheric pressure.

h. System 5, Sulfur Recovery and Tail Gas Treatment

Acid gas from Acid Gas Removal, System 4, is fed to a Claus-type three-stage sulfur recovery unit utilizing a proprietary process for handling lean H_2S acid gases. Typically, in a Claus-type sulfur plant, the acid gas is first passed through a knockout drum before entering the reaction furnace. The chemistry of the process involves converting H_2S to elemental sulfur according to the following equation:

$$2H_2S + SO_2 \rightarrow 3S + 2H_2O$$

The reactions are exothermic, and the heat liberated generates steam in the reaction furnace boiler and in the sulfur condenser. The sulfur from each condenser is drained to a recovery pit in By-Product Processing, System 13, and the tail gas from the final condenser is fed to a Beavon tail gas treating unit where essentially complete removal of the remaining sulfur compounds is achieved before discharge to the atmosphere. The Beavon sulfur removal process reduces the sulfur content in the tail gas to less than 100 ppm. In this system, hydrogenation and hydrolysis are used to convert essentially all sulfur compounds to hydrogen sulfide. This gas is then cooled and passed into a contactor where the hydrogen sulfide is absorbed by the redox solution and oxidized to elemental sulfur. The reduced redox solution is reoxidized by contact with air and subsequently recirculated to the contactor. Elemental sulfur is removed in the air-blowing step as a froth which is pumped to a sulfur melter to be melted under pressure, separated from the redox solution and transferred to By-Product Processing, System 13. The decanted redox solution is returned to the system.

The system receives 48,370 lbs/hr of acid gas from the Selexol solvent regenerator at about 7 psig and 120^{o}F.

The recovery of by-product elemental sulfur from the sulfur recovery system is 185 short tons/day per module, which amounts to an overall sulfur recovery of about 99.9 percent.

i. <u>System 13, By-Product Processing</u>

Sulfur is the only plant by-product other than ash/char solids which are disposed of on-site. Molten sulfur is pumped to a prilling tower in a continuously flowing circulation system. In the tower, sulfur is dispensed in droplets through nozzles. Droplets fall counter current to a stream of cooling air and solidify prior to landing in the bottom prill collection section. From the prill tower, sulfur is conveyed to a storage building from which it is transferred by truck or barge for sale.

j. System 14, Plant Power System

This system is generally designed to receive medium voltage electrical power (4.16 KV, 6.9 KV or 13.8 KV) and provide the following functions:

(1) Develop the necessary voltage stepdown arrangement for plant requirements

(2) Distribute the necessary power to the plant equipment.

TVA's incoming substation transformers receive power from its prevalent distributed voltage switching station and step down this voltage to a medium voltage to supply the plant electrical power requirement for motors, heaters, lighting, and other miscellaneous loads.

The Medium Voltage Electrical Distribution Systems is a secondary selection system (double ended supply) with several medium voltage buses. Each medium voltage bus receives power from its respective incoming substation transformer through an incoming breaker and supplies power to the medium voltage distribution system through the feeder breakers.

The Low Voltage Electrical Distribution System typically consists of multiple 480 V double-ended load centers and 480 V motor control centers (MCC's) supplying the power to 480 V loads throughout the plant. Two load centers are interconnected through a normally open tie breaker. In the event of loss of one load center transformer or its feeder, the 480 V loads of the affected load center are fed by the second load center through the tie breaker.

Each load center consists of an incoming line section, load center transformer, and low voltage section with metal enclosed draw out power circuit breakers.

Load center transformers are air cooled, dry type, 150^{O}F temperature rise, with delta connected primaries and wye connected secondaries. All load center feeder circuit breakers are 1600A frame and 50,000A RMS symmetrical interrupting capacity. The 480 V motor feeder breakers are electrically operated with instantaneous and long time trip units.

480 V MCC's consist of starters, feeder circuit breakers and control devices, assembled in a common structure with horizontal and vertical buses.

A 125 volt DC system supplies control power for medium voltage and 480 volt plant switchgear control, protective relaying and annunciation. The system also supplies power for emergency lighting.

k. System 15, Steam Generation/Distribution

Two of the largest steam demands within the plant are for the steam turbine drives of the air and oxygen compressors in the air separation plant. Since the coal gasifiers are the largest source of heat for generating steam within the plant, it was decided to produce 565 psia steam from the gasifiers and to drive the air and oxygen compressor turbines with that steam. In order to match the 565 psia steam supply (512,700 lbs/hr) with the horsepower demand of the air and oxygen compressors, it was necessary to superheat the steam to 1000^{O}F and to condense the turbine exhausts at 2.5" Hg and 109^{O}F.

The gasification section waste heat boilers were designed to produce 352,100 lbs/hr of 1500 psia steam to be used as follows:

Char recycle transport	32,700 lbs/hr
Sulfur plant (Claus gas reheaters)	20,000 lbs/hr
Raw gas compressor turbine	299,400 lbs/hr
Total	352,100 lbs/hr

The 299,400 lbs/hr of 1500 psia steam for the raw gas compressor turbine is superheated to 940^{O}F so that the turbine exhaust would produce 165 psia steam for driving the inert gas (nitrogen) compressor turbine and for supplying other users at that pressure or lower.

l. System 16, Water Supply

The raw water treatment unit is designed to provide treated and untreated water for the following facility water systems:

(1) Fire water
(2) Service water
(3) Potable water
(4) Cooling water
(5) Boiler feed water

Raw water is pumped from the river to a fire water-raw water storage tank.

The raw water-fire water storage tank provides surge capacity for water treatment as well as storage capacity for fire water. During an emergency, fire water is pumped from the tank to the fire water heater system. The fire water pumps are motor driven and have a diesel engine driven spare. The spare pump is equipped with automatic start-up capability in case of power failure.

The raw water is pumped from the raw water-fire water storage tank to the softener-clarifier. Lime, alum, and polyelectrolyte from the clarifier bulk chemical storage and feed system are added to the softener-clarifier, which is equipped with an internal flocculation mechanism. The alum and polyelectrolyte aid in the removal of suspended solids from the raw water. Lime is added during the clarification step to "cold soften" the raw water. Chlorine is added to the raw water to inhibit algae growth in the clarifier and sand filters and reduce organic contamination.

The underflow from the clarifier is a one wt percent sludge and is pumped to solids treatment for further processing.

The clarified and softened raw water from the softener-clarifier flows to the self-backwashing sand filters where additional suspended solids are removed. A pressure differential across the filter bed initiates the backwash cycle. The backwash flows by gravity to the sand filter backwash sump and is recycled to the softener-clarifier. The filtered water flows to the filtered water storage tank and is utilized as cooling tower make-up for the process cooling tower as service water for general plant use, as feed to the demineralizer package, and as feed to the potable water system.

Water intended for potable services is chlorinated and again filtered to meet American Water Works Association (AWWA) standards and stored in a tank sized to hold a day's potable water requirements. The chlorine residual is maintained at 0.5-1.0 ppm free chlorine in the tank.

Filtered water intended as feed to the demineralizer package is injected with sodium sulfide to remove trace amounts of chlorine which adversely affect the demineralizer resins and after filtered through activated carbon to remove any remaining organic contaminants and dissolve iron.

In the demineralizer, the mineral salts present in the water are removed by ion exchange. A two-step demineralization system, utilizing strong cation and strong anion exchangers in series, is provided. A degasifier following the strong cation and magnesium, which the anion exchangers remove anions such as chloride and sulfate. The strong anion exchanger also removes silica. The degasifier is provided to remove carbon dioxide and other dissolved gases.

The mixed bed polisher is provided to remove silica to 0.02 ppm and to polish returned turbine condensate for reuse.

The boiler feed water deaerating heaters operate at 30 psig and 250^{0}F. The deaerators reduce the oxygen content of BFW to 0.005 cc/liter.

Hydrazine or sodium sulfite is injected into the storage compartment of the deaerators for chemical scavenging of any residual oxygen. Morpholine is injected into the suction of the boiler feed water pumps to protect the condensate systems.

m. <u>System 17, Water Cooling</u>

The purpose of this unit is to provide cooling water to the various process users in the facility.

The cooling tower system includes the tower and fans, side stream filters, circulating water pumps, cold water basin, blowdown system, chemical addition equipment, and distribution system.

Cooling water is pumped from the cold water basin, through the distribution system to the process heat exchangers where low-level, sensible heat is picked up, and back to the cooling tower. The cooling tower rejects low-level heat by evaporative cooling to air drawn through the cooling tower by the cooling tower fans.

A portion of the circulating water is passed through side stream filters to reduce loading to suspended solids, dirt and scale.

The dissolved solids level of the cooling water is maintained by a continuous blowdown stream to the process condensate system. Water level in the cooling tower basis is maintained by continuous make-up of the clean water from the raw water treatment system.

The blowdown stream is passed through a blowdown treatment system to recover chromate ions via ion exchange or by chemical reduction to chromium hydroxide and is sent to waste treatment for disposal.

Chlorine is added to the cooling water on a routine periodic basis to prevent algae growth. Chemical algicides are added periodically to further eliminate algae growth. Sulfuric acid is added to control pH, and zinc and chromate inhibitors are added to the cooling water for corrosion control. Occasionally, a polyphosphate dispersant is added to enhance the action of the inhibitors.

n. System 18, Waste Water Treatment

The purpose of this unit is to collect and treat all plant liquid effluent streams. The plant design is predicated on "zero discharge" and permits recycle and reuse of treated water. Streams treated include the following:

(1) Oily water sewers
(2) Coal pile run-off
(3) Storm water run-off
(4) Demineralizer regenerant wastes and rinse water
(5) Cooling tower blowdown
(6) Sanitary waste water
(7) Gasifier slag quench drains

(8) Separated water from solids treatment

(9) Filtrate from biological treatment.

Process operations include:

(1) Oil Separator - streams containing free and dissolved oil and treated in a gravity separator utilizing an emulsion breaking chemical and heat to separate the oil-water mixture

(2) Sour Water Stripper - water streams with appreciable H_2O and HN_3 residuals are steam-stripped to remove these contaminants

(3) Equalization Basin - liquid streams with extremely high or low pH are mixed in an equalizing basis and treated with sulfuric acid or caustic to change the mixed pH to a value of 6.0 - 8.0

(4) Gravity Settling-Thickener - liquid streams with high suspended or dissolved solids are treated in a gravity settler-thickener and mixed with lime, alum, coagulant aids, and polymers to facilitate separation and thickening

(5) Multiple Effect Evaporation - neutralized wastes and brines are evaporated to recover water and concentrate the solids.

The recovered, treated water is used as make-up to cooling towers or raw water supply. The resultant solids are conveyed to the solids disposal system.

o. System 19, General Facilities

The purpose of this unit is to provide equipment or services to support the gasification facility at the facility level.

This unit is a general facility category and provides the following equipment and services:

(1) Administration building

(2) Laboratories

(3) Change rooms

(4) Warehouses

(5) Maintenance buildings

(6) Operation centers

(7) Security offices

(8) Plant air facility
(9) Fire house
(10) Visitor reception
(11) Plant fencing
(12) Plant lighting
(13) Roads, bridges
(14) Docking facilities
(15) Interconnection pipe ways
(16) Fire protection network
(17) Flare stacks and headers
(18) Plant instrument air compressors
(19) Environmental monitoring
(20) Site preparation.

H. LURGI BASED PLANT

A four module, 20,000 TPD, plant based on Lurgi coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five percent North Alabama coal. Medium Btu gas with heat content of 308 Btu/scf and not more than 200 ppm sulfur is the primary plant product. The plant is designed for zero water discharge. The design is based on 20,000 TPD as received coal feed to the gasifiers. Coal fines not processable in the Lurgi gasifiers are used to supplement tar and oil as fuel for steam generation. Excess is sold as a plant by-product. Thus total coal requirement exceeds 20,000 TPD in this case. Electricity from the TVA grid is used to supply the balance of the plant prime mover power requirements. The plant design was arrived at by reviewing the processes available for each system especially as they are catalogued in Appendix A. Based on the data available for these processes, a review of their suitability for the subject project and the design team experience, a block flow diagram was established and appropriate processes selected. The results of trade studies done in conjunction with other designs were incorporated as appropriate. Results of designs available in the literature were factored to meet the requirements of this project to provide material and energy balances as well as a basis for the cost analysis.

Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on steam factor.

The overall plant configuration is shown schematically in Figure IV.H.1. As shown, General Facilities, Instrumentation and Control, Coal Handling, and Solid Disposal serve plantwide functions. All other plant components are contained in four identical process modules. The instrumentation and control system operates to monitor overall plant performance and to control inter-module relationships. The total list of systems is given in Table IV.H.1.

The results of the design study are given in Table IV.H.2. Appendix B-4 contains the complete design report.

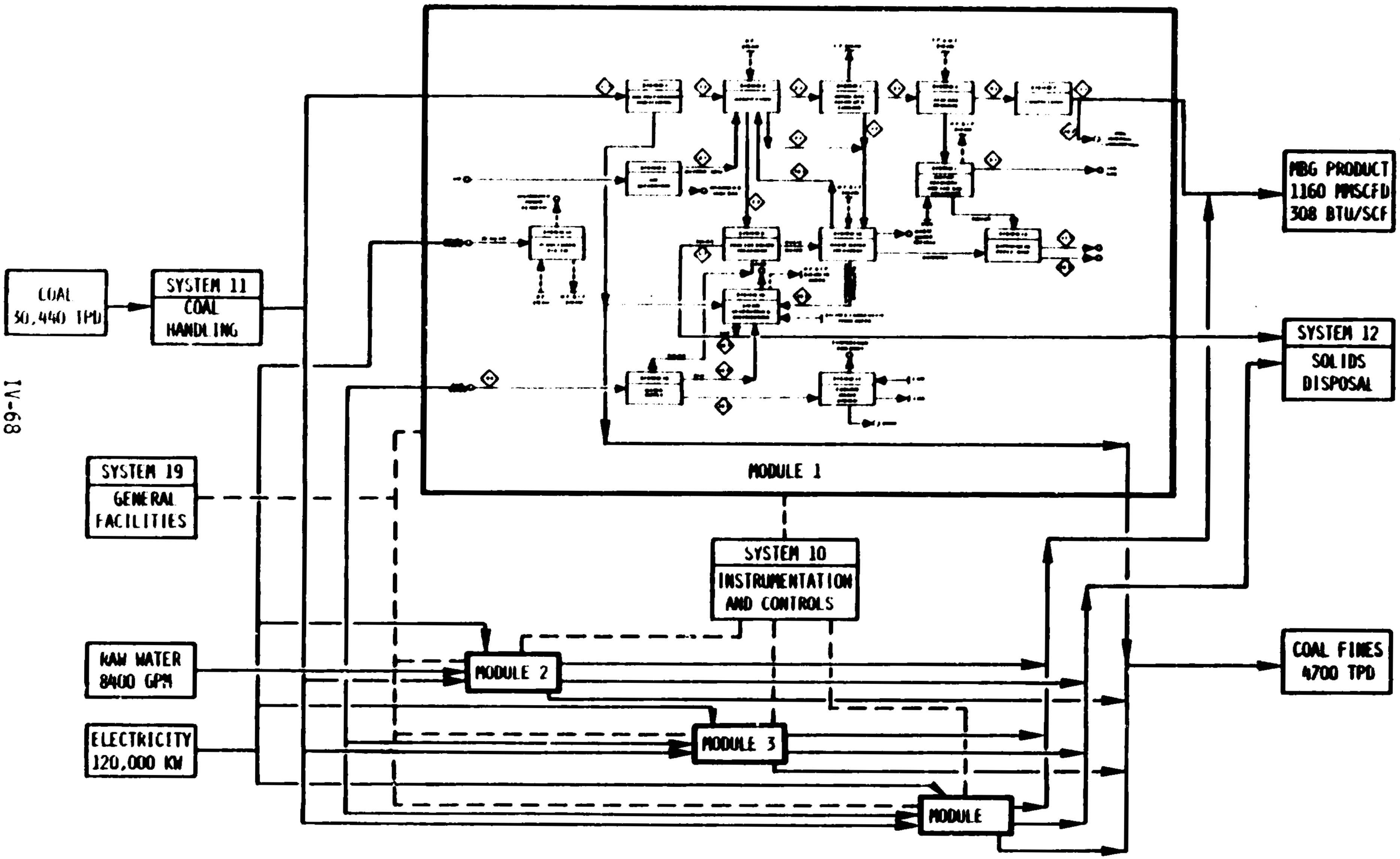

FIGURE IV.H.1 LURGI BLOCK FLOW DIAGRAM

TABLE IV.H.1. LIST OF SYSTEMS, LURGI PLANT

SYSTEM NO.	NUMBER OF PER MODULE	COST UNITS PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COAL PREPARATION & FEEDING
2	7	28	GASIFICATION
3	1	4	INITIAL GAS CLEANUP & COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	1	5	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10	-	1	INSTRUMENTATION AND CONTROL
11	-	1	COAL HANDLING
12	-	1	SOLIDS DISPOSAL
13	1	4	BYPRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	2	8	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKEUP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	-	1	GENERAL FACILITIES

TABLE IV.H.2(a). FACILITY PROCESS RESULTS SUMMARY

ITEM	LURGI
COAL, T/YR	10,000,000
FINES SOLD, T/YR	2,121,000
ELECTRICITY, 1000 Kwh/YR	934,000
YIELD MMSCFD	1,160
HHV BTU/SCF	308
COMPOSITION lb/hr	
Hydrogen	29,968
Nitrogen	3,928
Carbon Monoxide	152,127
Carbon Dioxide	363,892
Methane	45,720
Ethane	2,371
Light HC	3,892
Hydrogen Sulfide	61
Carbonyl Sulfide	300
Water	84
Total	602,343
SULFUR 1000 TONS/YR	58
AMMONIA 1000 TONS/YR	89
EFFICIENCY - PERCENT*	68.0

* EFFICIENCY BASED ON COAL CONSUMED, PURCHASED ELECTRICITY AS ELECTRICITY AND MBG PRODUCT.

TABLE IV.H.2(b). CONVERSION EFFICIENCY
LURGI PROCESS
20,000 TPD FACILITY

	10^6 BTU/HR	PERCENT
INPUTS		
1. COAL TO FACILITY	27,852	
2. ELECTRIC POWER TO FACILITY	100	
OUTPUTS		
3. MBG FROM FACILITY	14,912	
4. COAL FINES FROM FACILITY	5,908	
EFFICIENCY		
COAL-TO-MBG (3 ÷ 1) x 100%		53.5%
OVERALL PRODUCT EFFICIENCY 3 ÷ (1 + 2) x 100%		53.4%
OVERALL FACILITY EFFICIENCY 3 + 4 ÷ (1 + 2) x 100%		74.48%
GASIFICATION EFFICIENCY 3 ÷ (1 + 2 - 4)		68.0%

TABLE IV.H.2(c). OPERATING REQUIREMENTS FOR EXPECTED OPERATIONS LURGI PROCESS - PER MODULE

	BASIS	UNITS
Raw Materials		
Coal Import	TPY at 100% Operation	2,777,650 TPY
Coal Fines Export	TPY at 100% Operation	589,167 TPY
Catalyst and Chemical Makeup	100% Operation	1,204,500/YR
Utility Requirements		
Import Power	kWh/HR at 100% Operation	259,558,800 kWh/YR
Operating Requirements		
Labor		
Supervisors	Man-Hours/YR	30,000 mh/YR
Operators	Man-Hours/YR	145,720 mh/YR
Supplies	Factored as 15% of Operating Labor Cost	
Maintenance Requirements		
Labor	Factored at 1.6% of Total Depreciable Direct Investment	
Supplies	Factored at 2.4% of Total Depreciable Direct Investment	

1. Process Description

The plant consists of eighteen systems including General Facilities. However, this design was accomplished at the definition level only. There-fore, descriptions are given here only for the main process train systems. Other systems are generally similar in overall description to those given in Appendix A and Appendix B-4. A modular material balance is given in Figure IV.H.3.

a. System 1 - Coal Preparation and Feeding

This system receives 2"x0 coal from System 11, Coal Handling and crushes it to maximum 1 inch in size. Coal in the size range 1"x28 mesh is conveyed to the Coal Gasification Section. Minus 28 mesh coal is recovered and used as supplemental fuel in the steam generation and distribution system. Excess coal fines are sold as a plant by-product.

b. Coal Gasification

The Lurgi gasifier is a dry ash, gravitating bed type.

The gasifier is essentially a refractory-lined, water-jacketed cylindrical shell operating at 30 atmospheres pressure. Coal is received from Coal Preparation into lock hoppers situated above the gasification reactors. Coal is fed to the gasifier by gravitational feed from the lock hoppers and spread over the top of the bed of coal. The bed of coal gravitates from top to bottom. The coal flows counter-current to the gasification medium (oxygen and steam). Dry ash is removed continuously by a rotating grate into a semi-automatic ash lock.

The gas leaves the gasifier at a temperature of 650^{o}F. The gas is washed in a scrubbing cooler where it is cooled and water saturated. Traces of coal dust contained in the gas are removed via the action of heavy tar condensation on the particles in the scrubber. This mixture of tar-dust can be recycled to the gasifier. The scrubber is an integral part of the waste heat boiler system and the gas consists primarily of CO_2, CO, H_2 and CH_4, and some H_2S. The proportion of these components depend on the type of coal and the operating conditions. The gas also contains tar, oil, light naphtha, other hydrocarbons, and sulfur compounds.

TABLE IV.H.3. MATERIAL BALANCE LURGI PROCESS
5,000 TPD MODULE

(POUNDS PER HOUR)

	RAW COAL	GASIFIER FEED COAL	GASIFIER FEED STEAM	GASIFIER FEED OXYGEN	GASIFIER RAW GAS	COOLED RAW GAS	SWEETENED GAS	NET MBG PRODUCT
CARBON	386,083	253,635	-	-	-	-	-	-
HYDROGEN	27,282	17,925	-	-	30,030	30,030	30,030	29,968
OXYGEN	8,769	5,762	-	171,711	-	-	-	-
NITROGEN	36,378	23,902	-	3,068	3,937	3,937	3,937	3,928
SULFUR	23,514	15,449	-	-	-	-	-	-
CHLORINE	715	494	-	-	-	-	-	-
ASH	90,788	59,650	-	-	-	-	-	-
CARBON MONOXIDE	-	-	-	-	152,441	152,441	152,441	152,127
CARBON DIOXIDE	-	-	-	-	456,816	456,816	364,646	363,892
METHANE	-	-	-	-	45,815	45,815	45,815	45,720
ETHANE	-	-	-	-	2,376	2,376	2,376	2,371
LIGHT HC	-	-	-	-	3,899	3,899	3,899	3,892
TAR + OIL + NAPTHA	-	-	-	-	4,028	-	-	-
HCl	-	-	-	-	-	-	-	-
HYDROGEN SULFIDE	-	-	-	-	15,540	15,540	61	61
CARBONYL SULFIDE	-	-	-	-	660	660	300	300
AMMONIA	-	-	-	-	5,604	-	-	-
HYDROGEN CYANIDE	-	-	-	-	-	-	-	-
TOTAL DRY	573,515	376,817	-	174,779	721,146	711,514	603,505	602,259
WATER	60,652	39,850	441,392	-	869,881	1,342	312	84
TOTAL WET	634,167	416,667	441,392	174,779	1,591,027	712,856	603,817	602,343
T, ^{o}F	-	-	750	302	399	100	57	120
P, PSIA	-	-	665	515	440	430	427	615

c. System 3 - Initial Gas Cleaning and Cooling

Raw gas from the coal gasification system enters the initial gas cleaning and cooling system from the waste heat boiler in the coal gasification systems. Since this steam still contains ammonia, carbon dioxide, hydrogen sulfide, and oil and water vapor, special design considerations are required to prevent plugging and excessive fouling of cooling surfaces. This design including vertical tubes with tube walls washed with reinjected gas liquor is based on proprietary technology owned by Lurgi Kohle and Mineraloeltechnik, GmbH. Gas leaves this system at 100°F. Water effluent containing tar, oil, ammonia and acid gases is sent to Waste Water Treating.

d. Air Separation

Ninety-eight percent oxygen is provided at 515 psig. The oxygen requirement is 0.42 pound per pound as received coal fed to the gasifier.

e. System 7 - Compression

This system receives gas at 412 psig from the acid gas removal system and boosts the pressure to 600 psig for plant discharge.

f. System 15 - Steam Generation and Distribution

The steam generation and distribution system produces process steam from the gasifier jackets and high pressure steam from the waste heat boiler in the coal gasification system and medium and low pressure steam in the sulfur recovery system. Additional steam requirements are produced by high pressure steam boilers fired with tar, oil and phenolic liquids supplemented with coal fines from the coal preparation and feed system.

g. System 18 - Waste Water Treating

The waste water treatment system is designed for tar/oil separation, process condensate treatment, phenol recovery, ammonia recovery, and return of treated water to the plant. Solids removed in this system are put in the solids disposal system.

Tar/oil separation is accomplished in gravity settling tanks. The tar oil product is used as fuel in the steam generation and distribution system. Phenolic compounds are recovered in a Phenosolvan unit. Phenolic compounds also are burned as fuel pending develop- ment of a market. Ammonia is recovered in a Phosom-W unit and sent to the by-product process system for sale.

Streams containing miscellaneous free and dissolved oil are treated in a gravity separator utilizing an emulsion breaker and heat to separate the oil water mixture.

Streams containing a high or low pH are treated with sulfuric acid or lime as appropriate in equalization basins. Neutralized brines are evaporated with the concentrated solids going to solids disposal. Recovered treated water is used as makeup for the cooling tower and raw water supply.

I. BGC/LURGI BASED PLANT

A four module, 20,000 TPD, plant based on Lurgi coal gasification technology has been designed. The plant processes Kentucky No. 9 coal with provisions for up to five percent North Alabama coal. Medium Btu gas with heat content of 384 Btu/scf and not more than 200 ppm sulfur is the primary plant product. The plant is designed for zero water discharge. The design is based on 20,000 TPD as received coal feed to the gasifiers. Coal fines processable in the Lurgi gasifiers are used to supplement tar and oil as fuel for steam generation. Excess is sold as a plant by-product. Thus total coal requirement exceeds 20,000 TPD in this case. Electricity from the TVA grid is used to supply the balance of the plant prime mover power requirements. The plant design was arrived at by reviewing the processes available for each system especially as they are catalogued in Appendix A. Based on the data available for these processes, a review of their suitability for the subject project and the design team experience, a block flow diagram was established and appropriate processes selected. The results of trade studies done in conjunction with other designs were incorporated as appropriate. Results of designs available in the literature were factored to meet the requirements of this project to provide material and energy balances as well as a basis for the cost analysis.

Cost studies assumed a staggered construction schedule for the four modules beginning spring 1981 and a 90 percent on steam factor.

The overall plant configuration is shown schematically in Figure IV.I.1. As shown, General Facilities, Instrumentation and Control, Coal Handling, and Solid Disposal serve plantwide functions. All other plant components are contained in four identical process modules. The instrumentation and control system operates to monitor overall plant performance and to control inter-module relationships. The total list of systems is given in Table IV.I.1.

The results of the design study are given in Table IV.I.2. Appendix B-4 contains the complete design report.

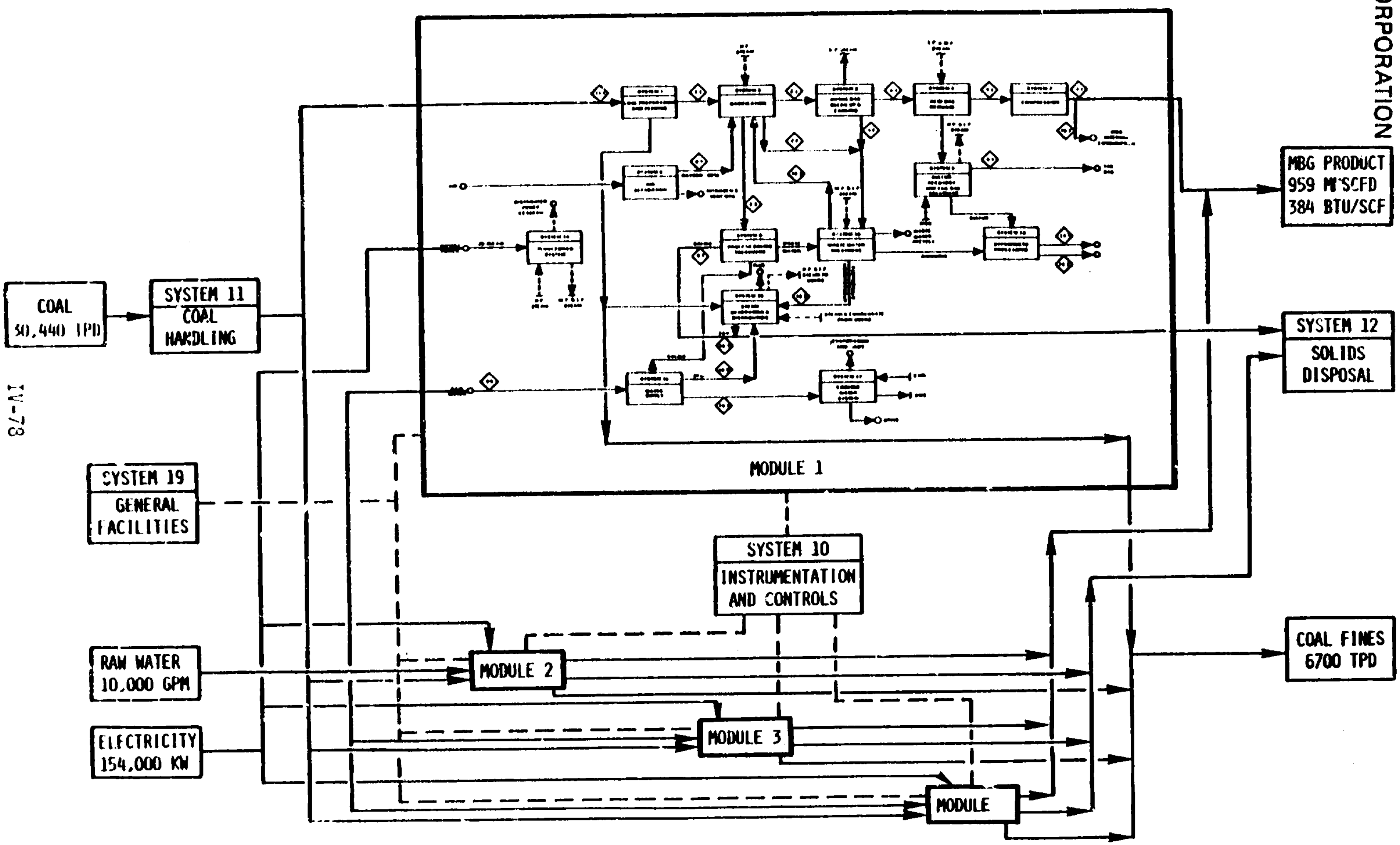

FIGURE IV.I.1 BGC/LURGI BLOCK FLOW DIAGRAM

TABLE IV.I.1. LIST OF SYSTEMS, BGC/LURGI

SYSTEM NO.	NUMBER OF COST UNITS PER MODULE	PER FACILITY	SYSTEM DESCRIPTION
1	1	4	COAL PREPARATION & FEEDING
2	3	12	GASIFICATION
3	1	4	INITIAL GAS CLEANUP & COOLING
4	1	4	ACID GAS REMOVAL
5	1	5	SULFUR RECOVERY
6	1	5	AIR SEPARATION
7	1	4	COMPRESSION
8	1	4	PROCESS SOLIDS TREATMENT
10		1	INSTRUMENTATION AND CONTROL
11	-	1	COAL HANDLING
12	-	1	SOLIDS DISPOSAL
13	1	4	BYPRODUCT PROCESSING
14	1	4	PLANT POWER SYSTEM
15	2	8	STEAM GENERATION/DISTRIBUTION
16	1	4	RAW WATER MAKEUP
17	1	4	COOLING WATER SYSTEM
18	1	4	WASTE WATER TREATMENT
19	-	1	GENERAL FACILITIES

TABLE IV.I.2(a). FACILITY PROCESS RESULTS SUMMARY

ITEM	BGC/LURGI
COAL, T/YR	10,000,000
INES SOLD, T/YR	3,918,000
ELECTRICITY, 1000 Kwh/YR	604,800
YIELD MMSCFD	960
HHV BTU/SCF	384
COMPOSITION lb/hr	
Hydrogen	15,183
Nitrogen	3,889
Carbon Monoxide	438,365
Carbon Dioxide	20,591
Methane	36,664
Ethane	1,901
Light HC	2,998
Hydrogen Sulfide	70
Carbonyl Sulfide	192
Water	70
Total	519,923
SULFUR 1000 TONS/YR	59
AMMONIA 1000 TONS/YR	89
EFFICIENCY - PERCENT*	76.4

* EFFICIENCY BASED ON COAL CONSUMED, PURCHASED ELECTRICITY AS ELECTRICITY AND MBG PRODUCT.

TABLE IV.I.2(b)
CONVERSION EFFICIENCY
BGC/SLAGGING LURGI PROCESS
20,000 TPD FACILITY

	10^6 BTU/HR	PERCENT
INPUTS		
1. COAL TO FACILITY	27,852	
2. ELECTRIC POWER TO FACILITY	525	
OUTPUTS		
3. MBG FROM FACILITY	15,268	
4. COAL FINES FROM FACILITY	8,404	
EFFICIENCY		
COAL-TO-MBG (3 ÷ 1) x 100%		54.8%
OVERALL PRODUCT EFFICIENCY 3 ÷ (1 + 2) x 100%		53.8%
OVERALL FACILITY EFFICIENCY 3 + 4 ÷ (1 + 2) x 100%		83.4%
GASIFICATION EFFICIENCY 3 ÷ (1 + 2 - 4) x 100%		76.4%

TABLE IV.I.2(c). OPERATING REQUIREMENTS FOR EXPECTED OPERATIONS LURGI/BGC PROCESS - PER MODULE

	BASIS	UNITS
Raw Materials		
Coal Import	TPY at 100% Operation	2,777,650 TPY
Coal Fines Export	TPY at 100% Operation	838,437 TPY
Catalyst and Chemical Makeup	$/YR at 100% Operation	$470,340/YR
Utility Requirements		
Import Power	kWh/HR at 100% Operation	168,000,000
Operating Requirements		
Labor		
Supervisors	Man-Hours/YR	27,091 mh/YR
Operators	Man-hours/YR	122,353 mh/YR
Supplies	Factored as 15% of Operating Labor Cost	
Maintenance Requirements		
Labor	Factored at 1.6% of Total Depreciable Direct Investment	
Supplies	Factored at 2.4% of Total Depreciable Direct Investment	

1. Process Description

The plant consists of eighteen systems including General Facilities. However, this design was accomplished at the definition level only. There-fore, descriptions are given here only for the main process train systems. Other systems are generally similar in overall description to those given in Appendix A and Appendix B-4. Table IV.I.3 contains the module material balance.

a. System 1 - Coal Preparation and Feeding

This system receives 2"x0 coal from System 11, Coal Handling and crushes it to maximum 1 inch in size. Coal in the size range 1"x28 mesh is conveyed to the Coal Gasification Section. Minus 28 mesh coal is recovered and used as supplemental fuel in the steam generation and distribution system. Excess coal fines are sold as a plant by-product.

b. Coal Gasification

This unit converts coal into medium heating value crude synthesis gas by partial oxidation in the presence of steam. The reaction takes place during countercurrent flow in a moving bed. The crude gas leaving the gasifier is scrubbed, quenched and saturated by gas liquor to remove coal dust and heavy tar. This multi-phase stream enters a waste heat exchanger for cooling and further condensation of heavier hydrocarbons prior to further processing.

The coal and flux are bed to the coal bunkers by a belt conveyor system. The feed chutes at the bottom of the coal bunkers control the flow of coal into the coal locks. Each gasifier has two coal locks that operate automatically on a cyclic basis. There coal locks are pressurized with an inert gas and feed alternately the coal surge vessel.

In the process, the coal and flux entering the top of the gasifier descends in a moving bed in countercurrent flow to the steam, oxygen and produced gas. While traveling from the top to the bottom of the gasifier, the coal is dried, devolatilized, and gasified. The heat

TABLE IV.I.3. MATERIAL BALANCE
BGC/SLAGGING LURGI PROCESS
5,000 TPD MODULE

(POUNDS PER HOUR)

	RAW COAL	GASIFIER FEED COAL	GASIFIER FEED STEAM	GASIFIER FEED OXYGEN	GASIFIER RAW GAS	COOLED RAW GAS	SWEETENED GAS	NET MBG PRODUCT
CARBON	386,032	253,635	-	-	-	-	-	-
HYDROGEN	27,282	17,925	-	-	15,220	15,200	15,214	15,183
OXYGEN	36,379	23,902	-	169,519	-	-	-	-
NITROGEN	8,770	5,762	-	3,029	3,898	3,898	3,898	3,889
SULFUR	23,515	15,450	-	-	-	-	-	-
CHLORINE	752	494	-	-	-	-	-	-
ASH	90,787	59,650	-	-	-	-	-	-
CARBON MONOXIDE	-	-	-	-	440,023	440,023	439,242	438,365
CARBON DIOXIDE	-	-	-	-	38,403	38,403	20,632	20,591
METHANE	-	-	-	-	36,738	36,738	36,738	36,664
ETHANE	-	-	-	-	1,904	1,904	1,904	1,901
LIGHT HC	-	-	-	-	3,005	3,005	3,005	2,998
TAR + OIL + NAPTHA	-	-	-	-	4,028	-	-	-
HCL	-	-	-	-	-	-	-	-
HYDROGEN SULFIDE	-	-	-	-	15,377	15,377	70	70
CARBONYL SULFIDE	-	-	-	-	943	943	193	192
AMMONIA	-	-	-	-	5,604	-	-	-
HYDROGEN CYANIDE	-	-	-	-	-	-	-	-
TOTAL DRY	573,517	376,818	-	172,548	565,143	555,511	520,896	519,853
WATER	60,652	39,850	109,761	-	95,062	1,050	257	70
TOTAL WET	634,169	416,668	109,761	172,548	660,205	556,561	521,153	519,923
T, ^{o}F	-	-	750	302	299	100	57	120
P, PSIA	-	-	665	505	440	430	425	615

required for these three steps is supplied by the exothermic reaction between the carbon in the coal and the oxygen in the bottom of the gasifier. Flux is added to form a low melting temperature meteric.

After leaving the gasifier, the raw gas is scrubbed and cooled in the wash cooler scrubbers and the waste heat exchanger.

In the bottom of the gasifier, the coal ash melts as a eutectic with the added flux to form slag. The molten slag collects at the bottom and is tapped intermittently through a tap hole into the quench vessel. In the quench vessel, the slag granulates immediately upon contact with the quench water. The granulated slag falls into the slag hopper and is dumped intermittently to the solids treatment system.

The raw gas leaving the gasifier requires immediate treatment to remove as many impurities as possible. The initial treatment is provided in wash coolers operating in parallel. In the wash coolers, gas liquor is injected to quench and saturate the raw gas. A multi-phase stream containing raw gas, condensed hydrocarbons, dust, stream, and water flows from the wash coolers to the waste heat exchanger for additional treatment. The multi-phase flow enters the waste heat exchanger above the gas liquor level. The gas and liquor are separated, and the gas is cooled by producing low pressure steam. The gas liquor condensed from the raw gas is collected in a sump. A portion of the dusty gas liquor is utilized as quench water for the wash coolers with the remaining liquor being sent to the waste water treating system.

c. System 3 - Initial Gas Cleaning and Cooling

Raw gas from the coal gasification system enters the initial gas cleaning and cooling system from the waste heat boiler in the coal gas- ification systems. Since this steam still contains ammonia, carbon dioxide, hydrogen sulfide, and oil and water vapor, special design considerations are required to prevent plugging and excessive fouling of cooling surfaces. This design including vertical tubes with tube walls washed with reinjected gas liquor is based on

proprietary technology owned by Lurgi Kohle and Mineraloeltechnik, GmbH. Gas leaves this system at 100°F. Water effluent containing tar, oil, ammonia and acid gases is sent to Waste Water Treating.

d. Air Separation

Ninety-eight percent oxygen is provided at 515 psig. The oxygen requirement is 0.42 pound per pound as received coal fed to the gasifier.

e. System 4, Acid Gas Removal

A Selexol unit similar to those described in other designs was selected for use in this system prior to compression in System 7.

f. System 7, Compression

This system receives gas at 412 psig from the acid gas removal system and boosts the pressure to 600 psig for plant discharge.

g. System 15, Steam Generation and Distribution

The steam generation and distribution system produces process steam from the gasifier jackets and high pressure steam from the waste heat boiler in the coal gasification system and medium and low pressure steam in the sulfur recovery system. Additional steam requirements are produced by high pressure steam boilers fired with tar, oil and phenolic liquids supplemented with coal fines from the coal preparation and feed system.

h. System 18, Waste Water Treating

The waste water treatment system is designed for tar/oil separation, process condensate treatment, phenol recovery, ammonia recovery, and return of treated water to the plant. Solids removed in this system are put in the solids disposal system.

Tar/oil separation is accomplished in gravity settling tanks. The tar oil product is used as fuel in the steam generation and distribution system. Phenolic compounds are recovered in a Phenosolvan unit. Phenolic compounds are to also be burned as fuel pending development of a market. Ammonia is recovered in a Phosom-W unit and sent to the by-product process system for sale.

Streams containing miscellaneous free and dissolved oil are treated in a gravity separator utilizing an emulsion breaker and heat to separate the oil water mixture.

Streams containing a high or low pH are treated with sulfuric acid or lime as appropriate in equalization basins. Neutralized brines are evaporated with the concentrated solids going to solids disposal. Recovered treated water is used as makeup for the cooling tower and raw water supply.

CHAPTER V
COSTS

A. FACILITY COST SUMMARY

The cost of the five gasification processes analyzed in this study are compared in Table V.B.1.*

*Two cases are considered in the cost analysis of the B&W-based plant. In the first case, base equipment cost for System 2. Gasification, are multiplied by an installation factor of 2.31 to arrive at the installed cost. This factor was arrived at by back calculation from a more detailed cost analysis based on Koppers-Totzek technology as shown in Appendix D. In the second case, an installed equipment cost factor of 1.5 was used based on information from B&W and supplied to this study by NASA. In this report, the first case result is used followed by the second case result in parenthesis. It is noted that discussions presented in Chapter XI imply that higher capacity units such as B&W should have a lower installation factor than low capacity units.

The BGC-Lurgi process is the most cost-effective in terms of UAE cost of service of $11.54/MMBTU (current dollars) and product price of $4.31/MMBTU (constant 1980 dollars). The next most cost-effective system is Texaco, with UAE and product price values of $13.38 and $5.00 respectively. These values are 16 percent greater than the BGC-Lurgi values. The least cost-effective process is Koppers-Totzek, with a UAE of $17.79 and product price of $6.64, 54 percent greater than the values for BGC-Lurgi. Table V.B.2 lists the processes in order of cost-effectiveness and shows the product prices normalized to BGC-Lurgi.

The BGC-Lurgi process is lowest cost in both capital requirement and total O&M. The entries in Table V.B.1 show that BGC-Lurgi total facility investment (instant plant value) $1,387,000,000, and total

capit requirements, $2,061,000,000, are the lowest of all the processes. Total O&M, feedstock, catalysts and chemicals are $310,000,000 annually. Texaco, the second most cost-effective system, is almost identical in both capital and total O&M costs, but is significantly lower in annual product, producing 103 x 10^{12} BTU compared to 121 x 10^{12} BTU for BGC-Lurgi. This difference accounts for the 16 percent advantage of the BGC-Lurgi product price.

BGC-Lurgi has a low total O&M annual cost despite high feedstock, catalyst and chemical cost. The latter are $276,000,000 per year compared to Texaco, Koppers-Totzek, and Babcock and Wilcox identical values of $181,000,000 per year. The higher BGC-Lurgi feedstock, catalyst and chemical costs are offset by (1) a low O&M annual cost of $100,000,000 and (2) annual byproduct credits of $66,000,000.

The Lurgi (B&W) process ranks third in cost-effectiveness behind BGC-Lurgi and Texaco. This is due primarily to a significant difference in capital costs between BGC-Lurgi and Lurgi. The major contributors to the high cost of the Lurgi process are the wastewater treatment system, which is more than double the BGC-Lurgi, and steam generation and distribution, which is two-thirds greater for Lurgi than for BGC-Lurgi.

The low cost-effectiveness of K[illegible]tzek is driven by a combination of the highest total O&M annual costs, $[illegible],000,000, and the lowest annual product, 90 x 10^{12} BTU.

Detailed cost data for each process are found in Appendix D.

B. DATA BASE DEVELOPMENT

There were four subtasks performed to create the data base for design and analysis of alternative coal gasification facilities:

- Raw Materials Analysis
- Market Analysis for Byproducts
- Alternate Products Analysis

The results of these subtasks are given in detail in the Coal Gasification Catalog, Appendix A.

TABLE V.B.1. COST COMPARISON GASIFICATION PROCESSES
(Millions of dollars, unless otherwise noted)

Cost Category \ Process	Texaco	Koppers-Totzek	Babcock & Wilcox	BGC-Lurgi	Lurgi
Total Facility Investment (Instant Plant)	1,416	1,591	2,437	1,387	1,879
Total Capital Requirements	2,091	2,371	3,347	2,061	2,747
O&M, Feedstock, Catalyst and Chemicals *	310	370	319	310	366
Feedstock, catalyst and Chemicals	181	181	181	276	279
O&M	129	189	138	100	134
Byproduct credits	0	0	0	(66)	(47)
Annual Product (10^{12} BTU)	103	90	100	121	117
UAE Cost of Service Price (Current $/MM BTU)	$13.38	$17.79	$17.11(14.37)	$11.54	$14.56
Product Price (Constant 1980 $/MM BTU)	5.00	6.64	6.39 (5.37)	4.31	5.44

*Total facility and 90% service factor.

TABLE V.B.2. RANKING OF GASIFICATION PROCESSES BY COST-EFFECTIVENESS

Gasification Process	Normalized Product Price
BGC-Lurgi	1.00
Texaco	1.16
Lurgi	1.26
Babcock & Wilcox	1.48 (1.24)
Koppers-Totzek	1.54

1. System Characterization

The preparation of cost estimates and the performance of detailed process engineering calculations require that certain licensing or confidentiality agreements be in place with the owners of coal gasification and associated technologies in order to obtain information necessary to do this work.

In the absence of these agreements, the situation faced in this study, studies and evaluations obtained in the public domain can be used as a basis for preparing preliminary process designs and budget level factored estimates. These budget level factored estimates are quite often used in the Hydrocarbon and Chemical Processing Industries to make decisions on further spending on or investigation of specific projects.

Usually, A/E firms performing studies on gasification projects for clients in the public domain, such as EPRI, DOE, Bureau of Mines, etc., have the licensing agreements to perform detailed engineering and cost estimates. The data presented in these study reports are then summaries of the engineering and cost estimates which provide a scaling base when used with good engineering judgment.

a. Approach

The approach taken in preparing the system characterizations of this section has been to utilize the system designs and trade-off studies previously done by the BDM-Mittelahuser team as well as the other published and proprietary studies in the Team's libraries. These reference data have been reviewed and those references which are most applicable to the reference facility have been identified and summarized in Table V.B.3.

b. System Identification

NASA-MARSHALL has provided definition of the candidate system which comprise the integrated facility and have been characterized in this task. Table V.B.4 identifies these Candidate Systems.

TABLE V.B.3. REFERENCES

Ref. No.	Report No.	Gasifier Type	Product Gas	Plant Type	Coal	Date	A/E	Remarks
1.	EPRI AF-741	Koppers-Totzek	MBG	Liquefaction	Illinois #6	76	Parsons	
2.	EPRI AF-416	Lurgi, BGC Slagging, Foster Wheeler Entrained Flow		Combined Cycle	Illinois #6	75	Stone & Webster	
3.	FE 2447-13	Lurgi		Mobil Methanol to Gasoline & Fischer-Tropsch to produce gasoline, SNG, LPG	Sub-bituminous Wyoming	77	Mobil	
4.	PB 264 702	L, K-T, B&W, Winkler, Wellman-Galusha	LBG MBG	LBG, IBG Production for iron ore pelletizing	Lignite, Sub-bituminous, Bituminous	76	McKee	Limited cost information on K-T only
5.	EPRI AF-642	L, BGC Slagger, Foster-Wheeler, CE, T		Combined Cycle	Illinois #6, New Mexico	76	Fluor	
6.	EPRI AF-244	L, IGT, CE		Fuel Gas Production	Illinois #6	75	Fluor	
7.	EPRI AF-531	Lurgi, Koppers-Totzek	MBG	Retrofitting Power Plant	3.5% S coal	75	TVA	
8.	FE 1775-13	2-stage entrainment slagger		POGO (Power-Oil-Gas-Other) Design Facility	1) Eastern Interior Province 2) S. Appalachian Eastern Province 3) Powder River Rocky Mt. Province	77	Parsons	
9.	FE 1775-18	Modified Bi-Gas IGT Design	LBG, MBG SNG	MPOP (Multi-Process Demonstration Plant) gasification, combined-cycle, Syngas production, Fischer-Tropsch Indirect Liquefaction	Eastern Interior Province	77	Parsons	
10.	EPRI AF-916	Texaco	MBG	Combined Cycle	Illinois #6	76	Fluor	
11.	EPRI AF-880	Texaco			Low-sulfur Western	78	Parsons	
12.	FE 1775-7	Bi-Gas		Fischer-Tropsch Oil and SNG Production	Eastern Interior Province	77	Parsons	
13.	FE 2240-31	IGT SYGAS, BCR Bi-Gas, PERC Synthane, Lurgi		SNG Production	Pittsburgh #8	76	Braun	
14.	MLGW DEMO (Vol. I)	U-Gas	MBG	Produce Industrial Fuel Gas	High Sulfur Bituminous	79	Foster Wheeler	Limited System Information, No costs given
15.	GRACE DEMO	Texaco	MBG	Demo Plant		79	EBASCO	Flow sheets describe various systems and their components, but no cost information
16.	PROCON CCD	HYGAS	SNG	Demo Plant	Bituminous and Lignite	78	Procon	
17.	CONOCO/BGC CCD FE-2542-10 (Vol. 2)	BGC, Lurgi Slagger	SNG	Demo Plant	Illinois #6 (sensitivity of costs to Ohio 29 and Pittsburgh 8 given)	78	Foster-Wheeler	
18.	FE 2542-13					77	Continental Oil	Data base for Pipeline Demo Plant no cost information
19.	EPRI AF-753	Texaco	MBG	Combined Cycle	Illinois #6	76	Fluor	
20.	EPRI AF-791					-	Gibbs & Hill	Discusses preparation of coal for combustion and conversion. Gives equipment and cost information for coal preparation system for coals and gasifiers of interest.
21.	EPRI AF-615	Koppers-Totzek	MBG	Char Gasification using COED Process	Char W. Kentucky Pittsburgh	78	Koppers	Specific equipment and cost data not available
22.	EPRI AF-1227	Texaco	Methanol Fuel Oil		Illinois #6 Wyodak Sub-bituminous	77	Braun	
	EPA-600/7-78-01C	None	None	Coal-Fired Boiler With FGD	Eastern 4.8% Sulfur	78	A. D. Little	CEA Dual-alkali
23.	APPENDIX A REFERENCE CHARACTERIZATION							

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TABLE V.B.4. NASA SYSTEM IDENTITY

1. COAL PREPARATION AND FEEDING
2. GASIFICATION - KOPPERS-TOTZEK COAL GASIFICATION PROCESS (TEXACO)
3. INITIAL GAS CLEANUP & COOLING
4. ACID GAS REMOVAL
5. SULFUR RECOVERY AND TAIL GAS TREATMENT
6. AIR SEPARATION
7. COMPRESSION
8. PROCESS SOLIDS TREATMENT (DEWATERING)
9. INCINERATOR
10. INSTRUMENTATION AND CONTROL
11. COAL HANDLING
12. SOLIDS WASTE RECYCLING/DISPOSAL
13. BY-PRODUCT PROCESSING
14. PLANT POWER SYSTEM
15. STEAM GENERATION/DISTRIBUTION
16. WATER SUPPLY
17. WATER COOLING SYSTEM
18. WASTE WATER TREATMENT
19. GENERAL FACILITIES
20. ALTERNATE PRODUCTS

Data reported in the literature and studies has been based on a more discrete level of unit operations. In order to manipulate the reported data as little as possible, the BDM-Mittelhauser Team identified the pertinent unit operations that are typically included in a Coal Gasification Facility and obtained cost and system characterizations on that basis. Table V.C.1 identifies the unit operations that were used to summarize data for this report. The summarized data have been reported on the unit operation level.

c. System Description

A brief description of each unit operation was prepared and included in the Coal Gasification Data Catalog (Appendix A).

Each description addresses the cost and design drivers of the unit operations, as well as issues of critical technology.

Preparation of facility cost estimates and life cycle costing requires an estimation of facility operating and maintenance costs. The costs can be broken down on a unit operation basis, but they are usually estimated based on total capital investment of the facility. A common estimation practice is to represent the various costs elements in this category as a percentage of the equipment installed capital cost. Maintenance expense is usually estimated as 1 to 6 percent of capital investment with a 60/40 material to labor split.

For this reason, the references that were nost applicable to a coal gasification project were reviewed and the quantitative operating and maintenance costs were reported as a function of total installed capital.

2. Raw Material Analysis

An analysis of raw material requirements has been performed for each system that has been characterized to identify the type, quantity, quality, etc., of raw material (other than coal) required to support the TVA Coal Gasification Facility.

These raw material requirements have been described as to:

- Identity of raw material
- Quantity or consumption of each raw material
- Source
- Costs
- Availability
- Shipping requirements.

Appendix A summarizes the data.

3. Byproducts Market Analysis

A market analysis was performed for the byproducts from the TVA Coal Gasification Facility. Since the plant is not conceptually designed, and the by-products are dependent on process type and design, quantities of the by-products are not included in this report. The generic price, use, and future market expectations are as given in Appendix A for the byproducts considered in this analysis. Oxygen is a possible export because all modules will not become operational at the same time and all the oxygen produced will not be consumed until the entire facility becomes operational. Oxygen, nitrogen, argon and carbon dixoide are treated as gases, and transportation by pipeline or truck is not considered. Tar, slag, and ash contents are not defined, therefore they are treated generically. Sulphur is an excellent byproduct candidate for marketing.

Conclusions from the by-product market analysis are:

- Quantifying TVA region market penetration can only be done if quantity and specifications of byproduct are known.
- Gases will have to be liquefied or transported by pipeline over relatively short distances (liquefaction costs have not be determined).
- Most of the byproducts will probably be used by new industries locating near the Murphy Hill site.

- Not clear if TVA would sell byproducts directly to consumers, most probably would sell to distributors at facility fence.
- TVA design criteria specifies only truck and barge transportation, however rail costs are included.
- Slag and ash would probably be impounded unless environmental constraints on utilization are relaxed.
- Tar could be converted to fuel for industries during natural gas curtailment.
- Sulphur is a price byproduct candidate for marketing.
- Gases will probably have to be purified for market sales.

4. Alternate Product Analysis

Complete facility economic analysis requires the investigation of possible alternate products. Methane, methanol, hydrogen, and gasoline have been identified as possible candidates. The systems needed to produce these options have been characterized in the same manner as the MBG systems. Table V.B.5 is the system breakdown used to obtain cost and characterization data for the alternate product options.

The approach taken in characterizing the alternate products options was the same as was used in the MBG case. The primary difference was that the references available to obtain information were far less in this section. Table V.B.3 of this report contains the identification of references used in this subtask. One difference from the MBG case is that operation and maintenance (O&M) cost data were obtained directly rather than as a proportion of capital costs.

TABLE V.B.5. ALTERNATE PRODUCTS SYSTEM IDENTIFICATION

SYSTEM	
90	SHIFT CONVERSION
91	METHANATION
92	GAS DRYING
93	METHANOL SYNTHESIS
94	GASOLINE SYNTHESIS
95	HYDROGEN RECOVERY

C. SENSITIVITY ANALYSIS

1. Introduction

Effects on cost were analyzed for the sensitivity cases defined in Table V.C.1. The analysis was conducted for the Koppers-Totzek process as representative of all the processes.

2. Methodology

The same procedures were used to generate sensitivity results as were used in the analysis of the base case, with some exceptions:

- Capital Cost. The 50% increase was applied to the present value of capital, and the remaining computations of UAE and product price conducted as for the base case.
- Operating and Coal Costs. The variations in these two categories were applied to the present value of the respective subcategory of total O&M.
- Byproduct Value. Sulfur credits were used to reduce the present value of total O&M.
- Sulfur in Product Gas. The acid gas removal system capital and O&M costs were generated as input to the approach used for MBG base case costing.
- Product Gas Pressure. Compression system capital and O&M costs were scaled to account for the variation in pressure and the base case methodology followed using the new input.

All other cases, such as the economic evaluation factor variation, were reruns of the base case with direct modification to the parameters as input data.

3. Sensitivity Results

The results are summarized in Table V.C.2 for product price effects. The Table shows that:

- The greatest impact occurs when the economic factor is increased to 20%. This results in an increase of product price in constant 1980 dollars to $9.17 from the base case value of $6.64, an increase of 38.1%.

- The next most significant impact is due to service factor changes. At a 60% service factor, the product price increases by 23.3% to a value of $8.19. The increase accelerates as the service factor drops.
- The third most significant impact is the 50% coal cost increase, which raises the product price 18.3% to $7.86.
- A close fourth is the 50% increase in operating costs, producing a 15.9% increase in product price to $7.70.
- A capital costs increase of 25% has only half the impact of the operating cost increase. The resulting product price is $7.21, an 8.6% increase over the base case.

Small impacts of 6% or less are obtained from the variations due to sale of sulfur, changes in the design/construction period, changes in operating life, reduction of sulfur in the product gas, and variation in product gas pressure.

Detailed results are presented in Table V.C.3 to V.C.21. One result deserves special comment. The extension of operating life has opposite effects on UAE and product price as shown in Tables V.C.13 and V.C.14. The reason is that price escalation in the extended years is so great that 1980 prices have to drop to keep revenues from exceeding cost. By contrast, the UAE must rise to account for the increased present value of O&M costs.

TABLE V.C.1. SENSITIVITY ANALYSIS APPLIED TO COST OF GAS

		INCREMENT
1.	COAL COST	+ 50%
2.	CAPITAL COST VARIATION	+ 25%
3.	OPERATING COSTS	+ 50%
4.	SERVICE FACTORS (BASE CASE = 90%)	80%, 70%, 60%
5.	BYPRODUCT VALUE	SEE TABLE BELOW
6.	DESIGN/CONSTRUCTION PERIOD PER MODULE	± 1 YEAR
7.	OPERATING LIFE YEARS	+ 5, +10
8.	SULFUR IN PRODUCT GAS	TO 1.0 PPM
9.	PRODUCT GAS PRESSURE	MAX = 800 psi MIN = 200 psi 1/
10.	ECONOMIC EVALUATION FACTOR	T.B.D.

BY-PRODUCT VALUES FOR SENSIVITITY ANALYSIS 2/

SULFUR, $/TON	70.00
SULFURIC ACID, $/TON	60.00
AMMONIA (ANHYDROUS), $/TON	130.00
NAPHTHA (120-320°F), $/GAL	0.80
LIGHT OIL (300-700°F), $/GAL	0.80
TAR (700°F), $/GAL	0.60
PHENOLS, $/GAL	0.75
COAL FINES, $/TON	80% OF ROM COAL COST
EXPORT POWER, ¢/kWh	SAME AS COST TO PLANT
METHANOL, ¢/GAL	35

1/ LOWEST PRACTICAL VALUE ABOVE 200 psi PERMITTED BY DESIGN CONSTRAINTS (CONTRACTOR TO RECOMMEND VALUE).

2/ EXCEPT FOR COAL FINES AND ELECTRIC POWER, ESCALATE BYPRODUCT VALUES AT SAME RATE AS COAL PRICES.

TABLE V.C.2. SUMMARY OF SENSITIVITY RESULTS ON PRODUCT PRICE

CASE	PRODUCT PRICE (1980$/MMBTU)	VARIATION OF PRODUCT PRICE FROM BASE CASE (1980 $/MMBTU)	(%)
BASE CASE (90% Service Factor)	6.64	0	0
COAL COST INCREASE BY 50%	7.86	1.22	18.3
CAPITAL COST INCREASE BY 25%	7.21	.57	8.6
OPERATING COSTS INCREASE BY 50%	7.70	1.06	15.9
SERVICE FACTOR			
80%	7.03	.39	5.9
70%	7.53	.89	13.3
60%	8.19	1.55	23.3
SALE OF SULFUR BYPRODUCT AT $70/TON	6.43	-.21	-3.1
VARIATION IN DESIGN/ CONSTRUCTION PERIOD PER MODULE			
+ 1 YEAR	6.79	.15	2.3
- 1 YEAR	6.51	-.09	-2.0
VARIATION OF OPERATING LIFE			
+ 5 YEARS	6.40	-.24	-3.6
+10 YEARS	6.25	-.39	-5.9
REDUCE SULFUR IN PRODUCT GAS TO 1.0 PPS	6.81	.17	2.5
PRODUCT GAS PRESSURE (BASE CASE = 600 psi)			
200 psi	6.32	-.32	-4.9
800 psi	6.75	.11	1.6
ECONOMIC EVALUATION FACTOR (BASE CASE = 12%)			
8%	5.82	-.82	-12.3
16%	7.75	1.11	16.7
20%	9.17	2.53	38.1

TABLE V.C.3. CASE 1 - SENSITIVITY ANALYSIS - VARIATION OF COAL COST (+50%)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
FEEDSTOCK ANNUAL REQ. (MMBTU/YR)	36.07×10^6	36.07×10^6	-0-	MODULE	
COAL COST ($/MMBTU)	1.25	1.875	+50%	--	ASSUMPTION
ANNUAL FEED COST	45.087	67.630	+50%	MODULE	
TOTAL ANNUAL O&M	92.579	115.122	+24.4%	MODULE	
$O\&M_{PV}$ (FEEDSTOCK)	1901.54	2852.32	+50%	FACILITY	
$TOTAL_{PV}$ CAPITAL AND O&M	5670.39	6705.14	18.3%	FACILITY	
PRODUCTION (MMBTU/YR)	90,082,584	SAME	-0-	FACILITY	
PRODUCTION PRICE ($1980/MMBTU)	6.64	7.86	+18.3%	FACILITY	
UAE COST OF SERVICE ($/MMBTU)	17.79	21.04	+18.3%	FACILITY	

TABLE V.C.4. CASE 2 - SENSITIVITY ANALYSIS - VARIATION OF CAPITAL COST (+25%)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
DEPRECIABLE INVESTMENT PV	1900.26	2375.33	+25%	FACILITY	ASSUMPTION
NONDEPRECIABLE INVESTMENT $_{PV}$	53.31	66.64	+25%	FACILITY	ASSUMPTION
CAPITAL COSTS $_{PV}$	1953.57	2441.97	+25%	FACILITY	
TOTAL CAPITAL AND O&M $_{PV}$	5670.39	6158.79	+8.6%	FACILITY	
PRODUCTION (MMBTU/YR)	90,082,584	SAME	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	7.21	+8.6%	FACILITY	
COST OF SERVICE ($/MMBTU)	17.79	19.32	+8.6%	FACILITY	

TABLE V.C.5. CASE 3 - SENSITIVITY ANALYSIS - VARIATION OF OPERATING COSTS (+50%)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
TOTAL ANNUAL O&M (LESS FEEDSTOCK CAT & CHEM)	47.240	70.860	+50%	MODULE	ASSUMPTION
$O\&M_{PV}$ (LESS FEEDSTOCK CAT & CHEM)	1804.64	2706.96	+50%	FACILITY	
TOTAL O&M $COSTS_{PV}$	3716.82	4619.14	+24.3%	FACILITY	
TOTAL CAPITAL AND $O\&M_{PV}$	5670.39	6572.71	+15.9%	FACILITY	
PRODUCTION (MMBTU/YR)	90,082,584	SAME	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	7.70	+15.9%	FACILITY	
UAE COST OF SERVICE ($/MMBTU)	17.79	20.62	+15.9%	FACILITY	

TABLE V.C.5. CASE 4a - SENSITIVITY ANALYSIS - VARIATION OF SERVICE FACTOR (80% OF TOTAL PLANT OPERATING CAPACITY)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
OPERATING CAPACITY FACTOR	90%	80%	-11.1%	FACILITY	ASSUMPTION
FEEDSTOCK ANNUAL REQUIREMENT	36.07×10^6	32.06×10^6	-11.1%	MODULE	
COAL UNIT COST ($/MMBTU)	1.25	1.25	-0-	--	
ANNUAL FEEDSTOCK COST	45.087	40.077	-11.1%	MODULE	
CAT & CHEM MAKE-UP	.252	.224	11.1%	MODULE	
TOTAL FEEDSTOCK, CAT & CHEM	45.339	40.301	-11.1%	MODULE	
ELEC PWR REQ (KWH/YR)	836,906,670	743,919,040	-11.1%	MODULE	
ELEC PWR UNIT COST ($/KWH)	.027	.027	-0-	--	
ELEC PWR COST	22.596	20.086	-11.1%	MODULE	
WATER REQUIREMENT (GAL/YR)	1513.728×10^6	1345.536×10^6	-11.1%	MODULE	
WATER-UNIT COST ($/KGAL)	.80	.80	-0-	--	
WATER COST	1.211	1.076	-11.1%	MODULE	
ADMIN & GEN'L PLANT	2.250	2.117	- 5.9%	MODULE	
O&M (LESS FEED, CAT/CHEM)	47.238	44.463	- 5.9%	MODULE	
TOTAL O&M	92.577	84.764	- 8.4%	MODULE	
$O\&M_{PV}$ (FEEDSTOCK)	1901.54	1690.26	-11.1%	FACILITY	
$O\&M_{PV}$ (CAT&CHEM)	10.64	9.45	-11.1%	FACILITY	
SUBTOTAL	1912.18	1699.71	-11.1%	FACILITY	
ELEC PWR_{PV}	851.35	756.76	-11.1%	FACILITY	
WATER $COST_{PV}$	17.16	15.25	-11.1%	FACILITY	
ADMIN & GEN'L $PLANT_{PV}$	88.28	83.09	-5.9%	FACILITY	
$O\&M_{PV}$ (LESS FEEDSTOCK CAT & CHEM)	1804.64	1702.95	-5.6%	FACILITY	
TOTAL CAPITAL AND O&M $COSTS_{PV}$	5670.39	5333.02	-5.9%	FACILITY	
PRODUCTION (MMBTU/YR)	22,520,646	20,018,353	-11.1%	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	7.03	+ 5.9%	FACILITY	
UAE COST OF SERVICE PRICE ($/MMBTU)	17.79	18.82	+ 5.9%	FACILITY	

TABLE V.C.7. CASE 4b - SENSITIVITY ANALYSIS - VARIATION OF SERVICE FACTOR (70% OF TOTAL PLANT OPERATING CAPACITY)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
OPERATING CAPACITY FACTOR	90%	70%	-22.2%	FACILITY	ASSUMPTION
FEEDSTOCK ANNUAL REQUIREMENT	36.07×10^6	28.05×10^6	-22.2%	MODULE	
COAL UNIT COST ($/MMBTU)	1.25	1.25	-0-	--	
ANNUAL FEEDSTOCK COST	45.087	35.067	-22.2%	MODULE	
CAT & CHEM MAKE-UP	.252	.196	-22.2%	MODULE	
TOTAL FEEDSTOCK, CAT & CHEM	45.339	35.263	-22.2%	MODULE	
ELEC PWR REQ (KWH/YR)	836,906,670	650,927,410	-22.%	MODULE	
ELEC PWR UNIT COST ($/KWH)	.027	.027	-0-	--	
ELEC PWR COST	22.596	17.575	-22.2%	MODULE	
WATER REQUIREMENT (GAL/YR)	1513.728×10^6	1177.344×10^6	-22.2%	MODULE	
WATER-UNIT COST ($/KGAL)	.80	.80	-0-	--	
WATER COST	1.211	.942	-22.2%	MODULE	
ADMIN & GEN'L PLANT	2.250	1.985	-11.8%	MODULE	
O&M (LESS FEED, CAT/CHEM)	47.238	41.685	-11.8%	MODULE	
TOTAL O&M	92.577	76.949	-16.9%	MODULE	
$O\&M_{PV}$ (FEEDSTOCK)	1901.54	1478.98	-22.2%	FACILITY	
$O\&M_{PV}$ (CAT&CHEM)	10.64	8.27	-22.2%	FACILITY	
SUBTOTAL	1912.18	1487.25	-22.2%	FACILITY	
ELEC PWR_{PV}	851.35	662.16	-22.2%	FACILITY	
WATER $COST_{PV}$	17.16	13.35	-22.2%	FACILITY	
ADMIN & GEN'L $PLANT_{PV}$	88.28	77.90	-11.8%	FACILITY	
$O\&M_{PV}$ (LESS FEEDSTOCK CAT & CHEM)	1804.64	1601.26	-11.3%	FACILITY	
TOTAL CAPITAL AND O&M $COSTS_{PV}$	5670.39	4995.95	-11.9%	FACILITY	
PRODUCTION (MMBTU/YR)	22,520,646	17,516,058	-22.2%	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	7.53	+13.3%	FACILITY	
UAE COST OF SERVICE PRICE ($/MMBTU)	17.79	20.15	+13.3%	FACILITY	

TABLE V.C.8. CASE 4c - SENSITIVITY ANALYSIS - VARIATION OF SERVICE FACTOR (60% OF TOTAL PLANT OPERATING CAPACITY)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
OPERATING CAPACITY FACTOR	90%	60%	-33.3%	FACILITY	ASSUMPTION
FEEDSTOCK ANNUAL REQUIREMENT	$36.07X10^6$	$24.05X10^6$	-33.3%	MODULE	
COAL UNIT COST ($/MMBTU)	1.25	1.25	-0-	--	
ANNUAL FEEDSTOCK COST	45.087	30.058	-33.3%	MODULE	
CAT & CHEM MAKE-UP	.252	.168	-33.3%	MODULE	
TOTAL FEEDSTOCK, CAT & CHEM	45.339	30.226	-33.3%	MODULE	
ELEC PWR REQ (KWH/YR)	836,906,670	557,937,780	-33.3%	MODULE	
ELEC PWR UNIT COST ($/KWH)	.027	.027	-0-	--	
ELEC PWR COST	22.596	15.064	-33.3%	MODULE	
WATER REQUIREMENT (GAL/YR)	$1513.728X10^6$	$1009.152X10^6$	-33.3%	MODULE	
WATER-UNIT COST ($/KGAL)	.80	.80	-0-	--	
WATER COST	1.211	.807	-33.3%	MODULE	
ADMIN & GEN'L PLANT	2.250	1.835	-17.6%	MODULE	
O&M (LESS FEED, CAT/CHEM)	47.238	38.908	-17.6%	MODULE	
TOTAL O&M	92.577	69.133	-25.3%	MODULE	
$O\&M_{PV}$ (FEEDSTOCK)	1901.54	1267.70	-33.3%	FACILITY	
$O\&M_{PV}$ (CAT&CHEM)	10.64	7.09	-33.3%	FACILITY	
SUBTOTAL	1912.18	1274.79	-33.3%	FACILITY	
$ELEC\ PWR_{PV}$	851.35	567.57	-33.3%	FACILITY	
$WATER\ COST_{PV}$	17.16	11.44	-33.3%	FACILITY	
ADMIN & GEN'L $PLANT_{PV}$	88.28	72.71	-17.6%	FACILITY	
$O\&M_{PV}$ (LESS FEEDSTOCK CAT & CHEM)	1804.64	1499.57	-16.9%	FACILITY	
TOTAL CAPITAL AND O&M $COSTS_{PV}$	5670.39	4659.32	-17.8%	FACILITY	
PRODUCTION (MMBTU/YR)	22,520,646	15,013,764	-33.3%	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	8.19	+23.3%	FACILITY	
UAE COST OF SERVICE PRICE ($/MMBTU)	17.79	21.93	+23.3%	FACILITY	

TABLE V.C.9. CASE 9 - SENSITIVITY ANALYSIS - VARIATION OF PRODUCT GAS PRESSURE

KOPPERS-TOTZEK	BASE	SENSITIVITY 1	SENSITIVITY 2	
PRODUCT GAS PRESSURE (PSI)	600	200	800	
ACID GAS PRESSURE DROP	25	25	25	
GAGE PRESSURE	625	225	825	
GAGE TO ABSOLUTE CONV (ATMOS)	14.7	14.7	14.7	
COMPRESSION OUTLET PRESSURE (PSI)	639.7	239.7	839.7	
COMPRESSION INLET PRESSURE (PSI)	14.9	14.9	14.9	
COMPRESSION RATIO OUTLET/INLET	42.9	16.1	56.4	
K/(K-1) FOR K = 1.4	3.5	3.5	3.5	INVERSE = .28571
CR^((K-1)/K)	2.92691	2.21209	3.16489	
BHP FACTOR = K/(K-1) (CR^((K-1)/K) - 1)	6.74419	4.24232	7.57712	BHP1/BHP2 = BHP FACTOR1/BHP FACTOR2
BHP COMPRESSOR 1	38,500			
BHP COMPRESSOR 2	34,600			
BHP COMPRESSOR 3	21,300			
BHP COMPRESSION SYSTEM	94,400	59,381	106,059	
GHP = .9 BHP	84,960	53,443	95,453	
SLOPE DETERMINATION				
GAS HORSEPOWER (GHP)	1,000	3,900		
BASE COMPRESSOR COST (\$1970)	90,000	200,000		CENTRIFUGAL COMPRESSION
SLOPE (\$/GHP)	(200,000 - 90,000)/(3900 - 1000)	= 110,000/2,900	= 37.931	
BASE COST (X10^6)	3.275	2.079	3.673	COST2 = M(GHP2 - GHP1) + COST1
PRODUCT GAS PRESSURE (PSI)	600	200	800	
GHP	84,960	53,443	95,453	
COMPRESSOR BASE COST (X10^6)	3.275	2.079	3.673	
AUXILIARY SLOPE DETERMINATION				
GAS HORSEPOWER (GHP)	900	4,000		
AUXILIARY BASE COST (\$1970)	10,000	20,000		
SLOPE (\$/GHP)	(20,000 - 10,000)/(4,000 - 900)	= 10,000/3,100	= 3.226	
GHP	84,960	53,443	95,453	
AUXILIARY COST (\$1970) X 10^6	.281	.180	.315	AUX2 = M1(GHP2-GHP1) + AUX1
COMPRESSOR BASE COST (X10^6)	3.275	2.079	3.673	
TOTAL EQUIPMENT COST (\$1970)	3.556	2.259	3.988	
ESCALATION TO \$1980 (1.988)	7.069	4.491	7.928	
MODULAR COST (X2.15)	15.198	9.656	17.046	

TABLE V.C.10. CASE 5 - SENSITIVITY ANALYSIS - BYPRODUCT VALUES

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
SULFURIC ACID	-0-	-0-	-0-	--	
AMMONIA	-0-	-0-	-0-	--	
NAPHTHA	-0-	-0-	-0-	--	
LIGHT OIL	-0-	-0-	-0-	--	
TAR	-0-	-0-	-0-	--	
PHENOLS	-0-	-0-	-0-	--	
COAL FINES	-0-	-0-	-0-	--	
EXPORT POWER	-0-	-0-	-0-	--	
METHANOL	-0-	-0-	-0-	--	
SULFUR (TONS/YEAR)	-0-	60206	---	MODULE	15,273 LB/HR PER MODULE @ 100%
SULFUR VALUE ($/TON)	70.00	70.00	-0-	--	
ANNUAL SULFUR REVENUE	-0-	4.214	---	MODULE	@ 90%
GROSS ANNUAL O&M	47.240	47.240	-0-	MODULE	
NET ANNUAL O&M	47.240	43.026	-8.9%	MODULE	
SUBTOTAL FEEDSTOCK, CAT/CHEM	45.339	45.339	-0-	MODULE	
TOTAL FEEDSTOCK & CAT/CHEM, O&M	92.579	88.365	-4.6%	MODULE	
$O\&M_{PV}$ (LESS FEEDSTOCK, CAT/CHEM)	1804.64	1804.64	-0-	FACILITY	
FEEDSTOCK & $CAT/CHEM_{PV}$	1912.18	1912.18	-0-	FACILITY	
BYPRODUCT $CREDITS_{PV}$	-0-	177.73	---	FACILITY	
NET $O\&M_{PV}$	3716.82	3539.09	-4.8%	FACILITY	
CAPITAL $COST_{PV}$	1953.57	1953.57	-0-	FACILITY	
TOTAL CAPITAL AND $O\&M_{PV}$	5670.39	5492.66	-3.1%	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	6.43	-3.1%	FACILITY	
UAE COST OF SERVICE ($/MMBTU)	17.79	17.23	-3.1%	FACILITY	

TABLE V.C.11. CASE 6a - SENSITIVITY ANALYSIS - DESIGN/CONSTRUCTION PERIOD PER MODULE (+1 YEAR)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
CONSTRUCTION LIFE (MONTHS)	57	69		MODULE	ASSUMPTION
TOTAL FACILITY INVESTMENT	1591.101	1591.101	-0-	FACILITY	
OTHER CAPITALIZED COSTS	672.059	715.741	+6.5%	FACILITY	
LAND RELATED COSTS	2.694	2.694	-0-	FACILITY	
SUB. DEPRECIABLE INVEST.	2265.854	2309.536	+1.9%	FACILITY	
WORKING CAPITAL	76.901	76.901	-0-	FACILITY	
LAND	.900	.900	-0-	FACILITY	
SUB. NONDEPRECIABLE INVEST.	77.801	77.801	-0-	FACILITY	
TOTAL CAPITAL REQUIREMENTS	2343.655	2387.337	+1.9%	FACILITY	
TOTAL ANNUAL O&M COST	370.316	370.316	-0-	FACILITY	
FEEDSTOCK & CAT/CHEM$_{PV}$	1912.18	1836.70	-3.9%	FACILITY	
OTHER O&M COSTS$_{PV}$	1804.64	1728.95	-4.2%	FACILITY	
TOTAL O&M COSTS$_{PV}$	3716.82	3565.95	-4.1%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1917.33	+0.9%	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	51.61	-3.2%	FACILITY	
TOTAL CAPITAL COSTS$_{PV}$	1953.57	1968.94	+0.8%	FACILITY	
TOTAL CAPITAL AND O&M COSTS$_{PV}$	5670.39	5534.69	-2.4%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	6.79	+2.3%		
UAE PRODUCT PRICE ($/MMBTU)	17.79	19.45	+9.3%	FACILITY	

TABLE V.C.12. CASE 6b - SENSITIVITY ANALYSIS - DESIGN/CONSTRUCTION PERIOD PER MODULE (-1 YEAR)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
CONSTRUCTION LIFE (MONTHS)	57	45		MODULE	ASSUMPTION
TOTAL FACILITY INVESTMENT	1591.101	1591.101	-0-	FACILITY	
OTHER CAPITALIZED COSTS	672.059	629.642	-6.3%	FACILITY	
LAND RELATED COSTS	2.694	2.694	-0-	FACILITY	
SUB. DEPRECIABLE INVEST.	2265.854	2223.438	-1.9%	FACILITY	
WORKING CAPITAL	76.901	76.901	-0-	FACILITY	
LAND	.900	.900	-0-	FACILITY	
SUB. NONDEPRECIABLE INVEST.	77.801	77.801	-0-	FACILITY	
TOTAL CAPITAL REQUIREMENTS	2343.655	2301.239	-1.8%	FACILITY	
TOTAL ANNUAL O&M COST	370.316	370.316	-0-	FACILITY	
FEEDSTOCK & CAT/CHEM$_{PV}$	1912.18	1989.39	+4.0%	FACILITY	
OTHER O&M COSTS$_{PV}$	1804.64	1884.92	+4.4%	FACILITY	
TOTAL O&M COSTS$_{PV}$	3716.82	3874.31	+4.2%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1883.97	-0.9%	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	54.89	+3.0%	FACILITY	
TOTAL CAPITAL COSTS$_{PV}$	1953.57	1938.86	-0.8%	FACILITY	
TOTAL CAPITAL AND O&M COSTS$_{PV}$	5670.39	5813.17	+2.5%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	6.51	-2.0%		
UAE PRODUCT PRICE ($/MMBTU)	17.79	16.29	-8.4%	FACILITY	

TABLE V.C.13. CASE 7a - SENSITIVITY ANALYSIS - VARIATION OF OPERATING LIFE (+5 YEARS)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
OPERATING LIFE (YEARS)	22	27	+22.7%	FACILITY	ASSUMPTION
TOTAL CAPITAL REQUIREMENTS	2343.655	2343.655	-0-	FACILITY	
TOTAL ANNUAL O&M COST	370.316	370.316	-0-	FACILITY	
FEEDSTOCK & CAT/CHEM$_{PV}$	1912.18	2181.60	+14.1%	FACILITY	
OTHER O&M COSTS$_{PV}$	1804.64	2069.07	+14.7%	FACILITY	
TOTAL O&M COSTS$_{PV}$	3716.82	4250.67	+14.4%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1900.26	-0-	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	56.16	+5.3%	FACILITY	
TOTAL CAPITAL COSTS$_{PV}$	1953.57	1956.42	+0.1%	FACILITY	
TOTAL CAPITAL AND O&M COSTS$_{PV}$	5670.39	6207.09	+9.5%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	6.40	-3.6%	FACILITY	
UAE PRODUCT PRICE ($/MMBTU)	17.79	18.55	+4.3%	FACILITY	

TABLE V.C.14. CASE 7b - SENSITIVITY ANALYSIS - VARIATION OF OPERATING LIFE (+10 YEARS)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
OPERATING LIFE (YEARS)	22	32	+45.5%	FACILITY	ASSUMPTION
TOTAL CAPITAL REQUIREMENTS	2343.655	2343.655	-0-	FACILITY	
TOTAL ANNUAL O&M COST	370.316	370.316	-0-	FACILITY	
FEEDSTOCK & CAT/CHEM$_{PV}$	1912.18	2396.01	+25.3%	FACILITY	
OTHER O&M COSTS$_{PV}$	1804.64	22285.53	+26.6%	FACILITY	
TOTAL O&M COSTS$_{PV}$	3716.82	4681.54	+25.9%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1900.26	-0-	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	57.78	+ 8.4%	FACILITY	
TOTAL CAPITAL COSTS$_{PV}$	1953.57	1958.04	+ 0.2%	FACILITY	
TOTAL CAPITAL AND O&M COSTS$_{PV}$	5670.39	6639.58	+17.1%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/ MMBTU)	6.64	6.25	- 5.9%	FACILITY	
UAE PRODUCT PRICE ($/MMBTU)	17.79	19.32	+ 8.6%	FACILITY	

TABLE V.C.15. CASE 8 - SENSITIVITY ANALYSIS - SULFUR IN PRODUCT GAS (TO 1.0 PPM)

KOPPERS-TOTZEK	BASE CASE	SENSITIVITY	% CHANGE	LEVEL	
ACID GAS REMOVAL SYSTEMS	80.352	154.772	+92.6%	FACILITY	ASSUMPTION
ALL OTHER SYSTEMS	1223.915	1223.915	-0-	FACILITY	
TOTAL SYSTEM CAPITAL INV	1304.267	1378.687	+5.7%	FACILITY	
SUBTOTAL DEPRECIABLE INVESTMENT	2265.854	2378.477	+5.0%	FACILITY	
SUBTOTAL NONDEPRECIABLE INVESTMENT	77.801	78.797	+1.3%	FACILITY	
TOTAL CAPITAL REQUIREMENTS	2343.655	2457.275	+4.8%	FACILITY	
TOTAL ANNUAL O&M	370.316	374.588	+1.2%	FACILITY	
FEEDSTOCK & CAT/CHEM$_{PV}$	1912.18	1912.18	-0-	FACILITY	
OTHER O&M$_{PV}$	1804.64	1847.48	+2.4%	FACILITY	
BYPRODUCT REVENUES$_{PV}$	-0-	-0-	-0-	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1995.16	+5.0%	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	53.99	+1.3%	FACILITY	
TOTAL CAPITAL COST$_{PV}$	1953.57	2049.15	+4.9%	FACILITY	
CAPITAL AND O&M$_{PV}$	5670.39	5808.82	+2.5%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90082584	90082584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	6.81	+2.5%	FACILITY	
UAE PRODUCT PRICE ($/MMBTU)	17.79	18.23	+2.5%	FACILITY	

TABLE V.C.16. CASE 9 - SENSITIVITY ANALYSIS - VARIATION PRODUCT GAS PRESSURE (CONTINUED)

KOPPERS-TOTZEK	BASE	$SENSITIVITY_1$	$SENSITIVITY_2$	
PRODUCT GAS PRESSURE (PSI)	600	200	800	
APPLYING BASE SLOPE TO REFERENCE TOTAL SYSTEM COST				
GHP	84,960	53,443	95,453	
SLOPE ($/GHP)				37.931
REFERENCE EQUIP. COST ($X10^6$)	10.529	8.153	11.320	K-T DOCUMENT

$EC_1 = (1.988)(37.931)(53443 - 84960) + 10.529 \times 10^6 = [(-1.195)(1.988) + 10.529] \times 10^6 = (-2.376 + 10.529)10^6$

$EC_2 = (1.988)(37.931)(95453 - 84960) + 10.529 \times 10^6 = [(.398)(1.988) + 10.529] \times 10^6 = (.791 + 10.529)10^6$

	BASE	$SENSITIVITY_1$	$SENSITIVITY_2$
MODULAR FACTOR	2.15	2.15	2.15
TOTAL SYSTEM COST ($X10^6$)	22.638	17.529	24.338
PRODUCT GAS PRESSURE (PSI)	600	200	800
BHP	94,400	59,381	106,059
KW = BHP/1.341	70,395	44,281	79,089
BASE KW FACILITY	106,152.5		
NET KW FACILITY	35,757.5	35,757.5	35,757.5
ELEC POWER REQ @100% (KW)	106,152.5	80,038.5	114,846.5
KWH/YEAR	929,896,300	701,130,000	1,006,000,000

NOTES:

(1) THERE ARE OFFSETTING COSTS (+) AND (-) WHICH CAUSE THE ACID GAS REMOVAL SYSTEMS COST DIFFERENCES TO BE SMALL

(2) WE ARE AWARE OF DESIGNS OF THE SELEXOL ACID GAS REMOVAL SYSTEMS OPERATING AT 350 PSI. THE PRECISE LOWER PRESSURE LIMIT WHICH WILL PERMIT PRODUCT TO BE PRODUCED AT CURRENT SPECIFICATIONS WAS NOT DETERMINED IN THIS STUDY.

TABLE V.C.17. CASE 9a - SENSITIVITY ANALYSIS-VARIATION GAS PRESSURE (MIN 200 psi)

KOPPERS-TOTZEK	BASE	SENSITIVITY	% CHANGE	LEVEL	ASSUMPTION
PRODUCT GAS PRESSURE (psi)	600	200	-66.7%	FACILITY	ASSUMPTION
COMPRESSION SYSTEMS	90.552	70.116	-22.6%	FACILITY	
ACID GAS REMOVAL SYSTEMS	80.352	80.352	NEGLIGIBLE	FACILITY	
ALL OTHER SYSTEMS	1133.363	1133.363	-0-	FACILITY	
TOTAL SYSTEM CAPITAL INV.	1304.267	1283.831	-15.7%	FACILITY	
SUBTOTAL DEPRECIABLE INV.	2265.854	2212.194	-2.4%	FACILITY	
SUBTOTAL NONDEPRECIABLE INV.	77.801	74.329	-4.5%	FACILITY	
TOTAL CAPITAL REQUIREMENTS	2343.655	2286.523	-2.4%	FACILITY	
ELECTRIC POWER RQMTS (KWH/YR)	3.3476×10^9	2.524010^9	-24.6%	FACILITY	• 90% OPERATION
ELECTRIC POWER COSTS	90.384	68.148	-24.6%	FACILITY	• 90% OPERATION
TOTAL ANNUAL O&M	370.316	345.796	-6.6%	FACILITY	• 90% OPERATION
FEEDSTOCK & CAT/CHEM$_{PV}$	1912.18	1912.18	-0-	FACILITY	
OTHER O&M$_{PV}$	1804.64	1572.52	-12.9%	FACILITY	
BY-PRODUCT REVENUES$_{PV}$	-0-	-0-	-0-	FACILITY	
TOTAL O&M$_{PV}$	3716.82	3484.70	-6.2%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1856.08	-2.3%	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	50.94	-4.4%	FACILITY	
TOTAL CAPITAL COST$_{PV}$	1953.67	1907.02	-2.4%	FACILITY	
TOTAL CAPITAL AND O&M$_{PV}$	5670.39	5391.73	-4.9%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90.082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	6.32	-4.9%	FACILITY	
UAE PRODUCE PRICE ($/MMBTU)	17.79	16.92	-4.9%	FACILITY	

TABLE V.C.18. CASE 9b - SENSITIVITY ANALYSIS-VARIATION OF PRODUCT GAS PRESSURE (MAX 800)

KOPPERS-TOTZEK	BASE	SENSITIVITY	% CHANGE	LEVEL	
PRODUTE GAS PRESSURE (psi)	600	800	+33.3%	FACILITY	ASSUMPTION
COMPRESSION SYSTEMS	90.552	97.352	+7.5%	FACILITY	
ACID GAS REMOVAL SYSTEMS	80.352	80.350	NEGLIGIBLE	FACILITY	
ALL OTHER SYSTEMS	1133.363	1133.363	-0-	FACILITY	
TOTAL SYSTEM CAPITAL INV.	1304.267	1311.065	+0.5%	FACILITY	
SUBTOTAL DEPRECIABLE INV.	2265.854	228[illegible]8	+0.8%	FACILITY	
SUBTOTAL NONDEPRECIABLE INV.	77.801	76.[illegible]6	+0.3%	FACILITY	
TOTAL CAPITAL REQUIREMENTS	2343.655	2362.664	+0.8%	FACILITY	
ELECTRIC POWER RQMTS (KWH/YR)	3.3476×10^9	3.621[illegible]$\times 10^9$	+8.2%	FACILITY	@ 90% OPERATION
ELECTRIC POWER COSTS	90.384	97.784	+8.2%	FACILITY	@ 90% OPERATION
TOTAL ANNUAL O&M	370.316	378.472	+2.2%	FACILITY	@ 90% OPERATION
FEEDSTOCK & CAT/CHEM$_{PV}$	1912.18	1912.18	-0-	FACILITY	
OTHER O&M$_{PV}$	1804.64	1881.86	+4.3%	FACILITY	
BY-PRODUCT REVENUES$_{PV}$	-0-	-0-	-0-	FACILITY	
TOTAL O&M$_{PV}$	3716.82	3794.04	+2.1%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1914.95	+0.8%	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	54.10	+1.5%	FACILITY	
TOTAL CAPITAL COST$_{PV}$	1953.67	1969.05	+0.8%	FACILITY	
TOTAL CAPITAL AND O&M$_{PV}$	5670.39	5763.09	+1.6%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	6.75	+1.6%	FACILITY	
UAE PRODUCE PRICE ($/MMBTU)	17.79	18.08	+1.6%	FACILITY	

TABLE V.C.19. CASE 10a - SENSITIVITY ANALYSIS-ECONOMIC EVALUATION FACTOR-8%

KOPPERS-TOTZEK	BASE	SENSITIVITY	% CHANGE	LEVEL	
ECONOMIC EVALUATION FACTOR	12%	8%	-33%	FACILITY	ASSUMPTION
FEEDSTOCK$_{PV}$	1901.54	3447.52	+81%	FACILITY	
CATALYST & CHEMICALS$_{PV}$	10.64	19.28	+81%	FACILITY	
SUBTOTAL$_{PV}$	1912.18	3466.80	+81%	FACILITY	
ELECTRIC POWER$_{PV}$	851.35	1543.27	+81%	FACILITY	
WATER$_{PV}$	17.16	28.63	+66	FACILITY	
OPERATING LABOR$_{PV}$	71.17	129.02	+81%	FACILITY	
OPERATING SUPPLIES$_{PV}$	10.68	19.35	+81%	FACILITY	
MAINTENANCE LABOR$_{PV}$	256.38	464.71	+81%	FACILITY	
MAINTENANCE SUPPLIES$_{PV}$	384.57	697.07	+81%	FACILITY	
SUPERVISION$_{PV}$	22.13	40.12	+81%	FACILITY	
GENERAL PLANT$_{PV}$	102.92	186.57	+81%	FACILITY	
ADMIN & GENERAL$_{PV}$	88.28	160.03	+81%	FACILITY	
PROPERTY TAXES & INS.$_{PV}$	0.00	0.00	-0-	FACILITY	
SUBTOTAL$_{PV}$	1804.64	3268.78	+81%	FACILITY	
BYPRODUCT REVENUES$_{PV}$	0.00	0.00	-0-	FACILITY	
TOTAL O&M COSTS$_{PV}$	3716.82	6735.58	+81%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	2162.95	+13.8%	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	60.28	+13.1%	FACILITY	
TOTAL CAPITAL$_{PV}$	1953.57	2223.23	+13.8%	FACILITY	
CAPITAL AND O&M COSTS$_{PV}$	5670.39	8958.81	+58%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	5.82	-12.3%	FACILITY	
UAE PRODUCT PRICE ($/MMBTU)	17.79	16.84	-5.3%	FACILITY	

TABLE V.C.20. CASE 10b - SENSITIVITY ANALYSIS-ECONOMIC EVALUATION FACTOR-16%

KOPPERS-TOTZEK	BASE	SENSITIVITY	% CHANGE	LEVEL	
ECONOMIC EVALUATION FACTOR	12%	16%	+33%	FACILITY	ASSUMPTION
FEEDSTOCK$_{PV}$	1901.54	1115.40	-41.3%	FACILITY	
CATALYST & CHEMICALS$_{PV}$	10.64	6.24	-41.3%	FACILITY	
SUBTOTAL$_{PV}$	1912.18	1121.64	-41.3%	FACILITY	
ELECTRIC POWER$_{PV}$	851.35	499.97	-41.3%	FACILITY	
WATER$_{PV}$	17.16	10.83	-36.9%	FACILITY	
OPERATING LABOR$_{PV}$	71.17	41.77	-41.3%	FACILITY	
OPERATING SUPPLIES$_{PV}$	10.68	6.27	-41.3%	FACILITY	
MAINTENANCE LABOR$_{PV}$	256.38	150.49	-41.3%	FACILITY	
MAINTENANCE SUPPLIES$_{PV}$	384.57	225.73	-41.3%	FACILITY	
SUPERVISION$_{PV}$	22.13	12.99	-41.3%	FACILITY	
GENERAL PLANT$_{PV}$	102.92	60.41	-41.3%	FACILITY	
ADMIN & GENERAL$_{PV}$	88.26	51.81	-41.3%	FACILITY	
PROPERTY TAXES & INS.$_{PV}$	0.00	0.00	-0-	FACILITY	
SUBTOTAL$_{PV}$	1804.64	1060.27	-41.2%	FACILITY	
BYPRODUCT REVENUES$_{PV}$	0.00	0.00	-0-	FACILITY	
TOTAL O&M COSTS$_{PV}$	3716.82	2181.90	-41.3%	FACILITY	
DEPRECIABLE INVESTMENT$_{PV}$	1900.26	1675.43	-11.8%	FACILITY	
NONDEPRECIABLE INVESTMENT$_{PV}$	53.31	44.13	-17.2%	FACILITY	
TOTAL CAPITAL$_{PV}$	1953.57	1719.56	-12.0%	FACILITY	
CAPITAL AND O&M COSTS$_{PV}$	5670.39	3901.46	-31.2%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	7.75	+16.7%	FACILITY	
UAE PRODUCT PRICE ($/MMBTU)	17.79	19.41	+9.1%	FACILITY	

TABLE V.C.21. CASE 10c - SENSITIVITY ANALYSIS-ECONOMIC EVALUATION FACTOR-20%

KOPPERS-TOTZEK	BASE	SENSITIVITY	% CHANGE	LEVEL	
ECONOMIC EVALUATION FACTOR	12%	20%	+66.7%	FACILITY	ASSUMPTION
$FEEDSTOCK_{PV}$	1901.54	690.09	-63.7%	FACILITY	
CATALYST & $CHEMICALS_{PV}$	10.64	3.86	-63.7%	FACILITY	
$SUBTOTAL_{PV}$	1912.18	693.95	-63.7%	FACILITY	
ELECTRIC $POWER_{PV}$	851.35	309.94	-63.7%	FACILITY	
$WATER_{PV}$	17.16	7.14	-58.4%	FACILITY	
OPERATING $LABOR_{PV}$	71.17	25.87	-63.7%	FACILITY	
OPERATING $SUPPLIES_{PV}$	10.68	3.88	-63.7%	FACILITY	
MAINTENANCE $LABOR_{PV}$	256.38	93.20	-63.7%	FACILITY	
MAINTENANCE $SUPPLIES_{PV}$	384.57	139.80	-63.7%	FACILITY	
$SUPERVISION_{PV}$	22.13	8.04	-63.7%	FACILITY	
GENERAL $PLANT_{PV}$	102.92	37.41	-63.7%	FACILITY	
ADMIN & $GENERAL_{PV}$	88.28	32.09	-63.7%	FACILITY	
PROPERTY TAXES & $INS._{PV}$	0.00	0.00	-0-	FACILITY	
$SUBTOTAL_{PV}$	1804.64	657.37	-63.7%	FACILITY	
BYPRODUCT $REVENUES_{PV}$	0.00	0.00	-0-	FACILITY	
TOTAL O&M $COSTS_{PV}$	3716.82	1351.32	-63.7%	FACILITY	
DEPRECIABLE $INVESTMENT_{PV}$	1900.26	1482.64	-22.0%	FACILITY	
NONDEPRECIABLE $INVESTMENT_{PV}$	53.31	35.77	-32.9%	FACILITY	
TOTAL $CAPITAL_{PV}$	1953.57	1518.41	-22.3%	FACILITY	
CAPITAL AND O&M $COSTS_{PV}$	5670.39	2869.72	-49.4%	FACILITY	
ANNUAL PRODUCT (MMBTU)	90,082,584	90,082,584	-0-	FACILITY	
PRODUCT PRICE ($1980/MMBTU)	6.64	9.17	+38.1%	FACILITY	
UAE PRODUCT PRICE ($/MMBTU)	17.79	21.69	+21.9%	FACILITY	

CHAPTER VI
ALTERNATE PRODUCTS ANALYSIS

A. PURPOSE AND APPROACH

The purpose of this analysis is to provide cost estimates for potential a'ternative products to aid in product mix and process technology decisions for the facilities. Designs were developed at two levels, preliminary and definitive. Preliminary designs and cost estimates were developed by factoring flows and cost versus capacity from representative systems in previously published designs. The definitive designs were prepared in accordance with the conceptual design methodology described in Chapter IV.

The preliminary designs were developed as "add-on" modules, i.e., as separate plants receiving MBG "over the fence," produced to TVA specifications. This approach was based on the assumption that alternate product production would function as a temporary load leveler while the demand for MBG grows to equal plant capacity. The product costs, however, are based on the assumption that the alternate product modules are operated at 90 percent of design capacity for 20 years, the life of the MBG module. The definitive design were developed as fully integrated plant. Three sets of cases were developed as follows:

I. Koppers-Totzek and Texaco Single Product Facilities

Koppers-Totzek	to methane	:	preliminary
Koppers-Totzek	to methanol	:	preliminary
Koppers-Totzek	to gasoline	:	preliminary
Koppers-Totzek	to hydrogen	:	preliminary
Texaco	to methane	:	preliminary
Texaco	to methanol	:	preliminary
Texaco	to gasoline	:	preliminary
Texaco	to hydrogen	:	preliminary

II. Lurgi Single Product Facilities

Lurgi	to methane	:	preliminary
Lurgi	to methanol/methane	:	preliminary

III. Mixed Product Facilities

Koppers-Totzek and Texaco to MBG and methane	:	definitive
Koppers-Totzek to MBG and methane	:	definitive

The cost results for each set are discussed in the following sections, and the designs are briefly summarized. A detailed discussion of the designs and associated analyses and tradeoffs that led to specific process selections is presented in Appendix G.

The potential marketability of the alternate products is discussed in Chapter VIII.

B. KOPPERS-TOTZEK AND TEXACO SINGLE PRODUCT FACILITIES

Add-on modules were designed for methane, methanol, gasoline, and hydrogen for both the Koppers-Totzek and Texaco gasifiers. The total plant consists of four add-on modules appended to the MBG plants as described in Chapter II. For the cost evaluation it is assumed that the plant produces only the alternate product over the entire life of the plant, with a 90% on-stream factor.

1. Cost Evaluation

In every instance, the Texaco products are less costly than the Koppers-Totzek products. This is due to the considerably higher efficiency of the Texaco gasifier, as evidenced in the higher product yield. The higher product yield and lower operating cost of the Texaco gasifier more than compensate for its higher capital cost.

The cost of methane, methanol, and hydrogen per million BTU are approximately equal (hydrogen is somewhat higher for Koppers-Totzek) with gasoline being about 20% higher. Capital cost contributes about

	METHANE		METHANOL		GASOLINE		HYDROGEN	
	KT	TEXACO	KT	TEXACO	KT	TEXACO	KT	TEXACO
TOTAL CAPITAL INVESTMENT*	2411	2827	2400	2816	2760	3130	2926	3547
O&M, FEED*	345	335	373	365	380	374	350	341
PRODUCT YIELD 10^{12} BTU/YEAR	76	84	79	89	61	79	75	95
PRICE								
$1980/MMBTU	8.03	7.63	8.08	7.54	11.21	9.04	8.94	7.61
$1980/GALLON	-	-	0.53	0.48	1.25	1.01	-	-
$1980/MMSCF	8.03	7.63	-	-	-	-	2.90	2.47

NOTE: 1) Costs are for total plant, based on a single module design with add-on facility.
2) All costs are in millions of dollars.

* Instant Plant

Figure VI.B.1. Koppers-Totzek and Texaco Single Product Plant Costs

40% of the price, and operations and maintenance and fuel about 60%. As a fraction of price, Koppers-Totzek operating costs are higher than Texaco, due to Koppers-Totzek's lower efficiency. Consequently, Koppers-Totzek is more vulnerable than Texaco to unanticipated price increases in coal.

2. Description of Designs

a. Introduction

A simplified block flow diagram is presented for each alternate product add-on facility. Design details and analyses are provided in Appendix G. The first four cases represent add on modules for methane, methanol, gasoline, and hydrogen. Although the diagrams shown are for Koppers-Totzek, they also apply to the corresponding Texaco cases, except that Texaco does not require compression for methane.

b. Methane

The objective of this design is to define the process and costs related to the production of pipeline-quality (high BTU) gas. The input to the process is MBG consisting primarily of H_2, CO, CO_2, H_2O, H_2S and COS. The output is fully interchangeable high BTU gas or methane.

The conventional approach is to shift the CO to CO_2, remove the acid gas, and then perform methanation. The advantage here is that proven technology can be used. The disadvantage is that water is condensed from the process stream twice, requiring extra steam.

The final approach selected is to modify the conventional approach such that acid gas removal proceeds shift followed by methanation and a final CO_2 removal. The advantage is the same as for the combined shift/methanation scheme discussed above, with the added benefit of being able to use proven equipment. The advantage of this approach is that the steam injected ahead of the shift stays in the gas. However, an extra CO_2 removal step is required.

Compression and drying is required to meet final pipeline specifications. Drying is a low cost item and needs to be the last process in the overall system. However, the position of the compressor(s)

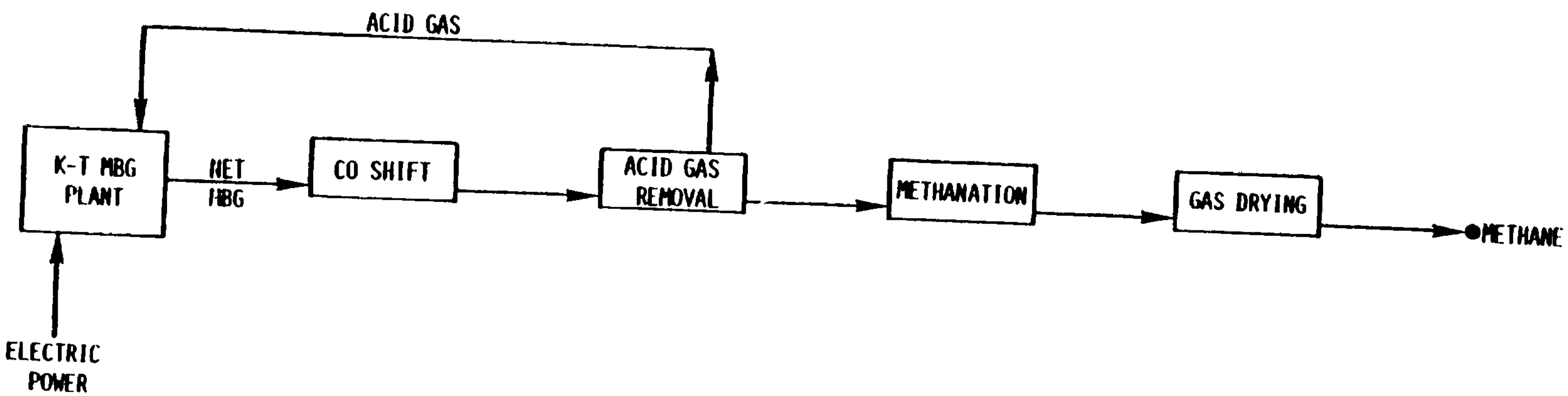


Figure VI.C.1. Koppers-Totzek to Methane

in the system must be determined and several locations are possible. Compression prior to acid gas removal lowers the AGR cost, but requires the compression of the sulfur gases and the CO_2. Compression between the AGR and methanation lowers methanation costs and there is less gas to compress after CO_2 and sulfur gas removal. However, it is still necessary to compress H_2 and CO rather than the CH_4. Compression after methanation requires that only one-half the volume of gas be compressed. Based on these factors, it was decided to compress to pipeline pressure of 1000 psig after methanation, since the synthesis gas is available at 600 psig.

c. Methanol

The design objective for methanol production is to produce a fuel-grade methanol with a purity of more than 95% (wt.) from MBG consisting primarily of H_2, CO, CO_2, H_2O, H_2S, COS. Extra high purity (greater than 99% wt.) is not required.

There are two process step alternatives for a methanol synthesis system. The first is AGR followed by shift, CO_2 removal and methanol synthesis. The second is a shift followed by AGR and then methanol synthesis. An evaluation of these alternatives indicated that the additional cost for another CO_2 removal step in option 1 made option 2 more attractive. Therefore, the second alternative was chosen.

d. Gasoline

The objective of this design is to produce a motor quality gasoline while creating a minimum of byproducts that have maximum marketability.

There are two alternate process-step sequences available. The Mobil M process dehydrates methanol. The second alternate, Fischer-Tropsch synthesis, is direct synthesis of CO and H_2 to a spectrum of liquids. Fischer-Tropsch synthesis requires upgrading of the reaction products by means of one or more of the refinery-type processes, including hydrotreating, isomerization, reforming, and polymerization.

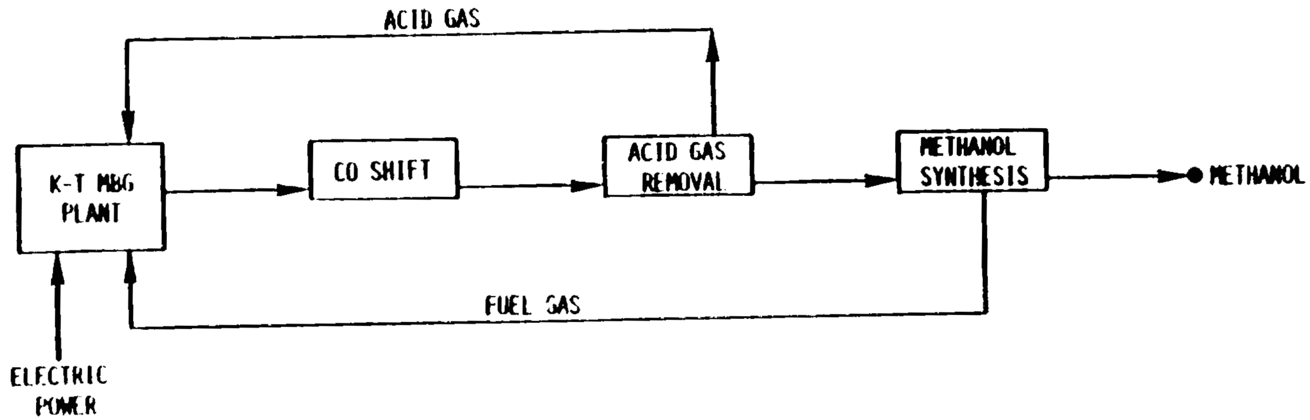


Figure VI.C.2. Koppers-Totzek to Methanol

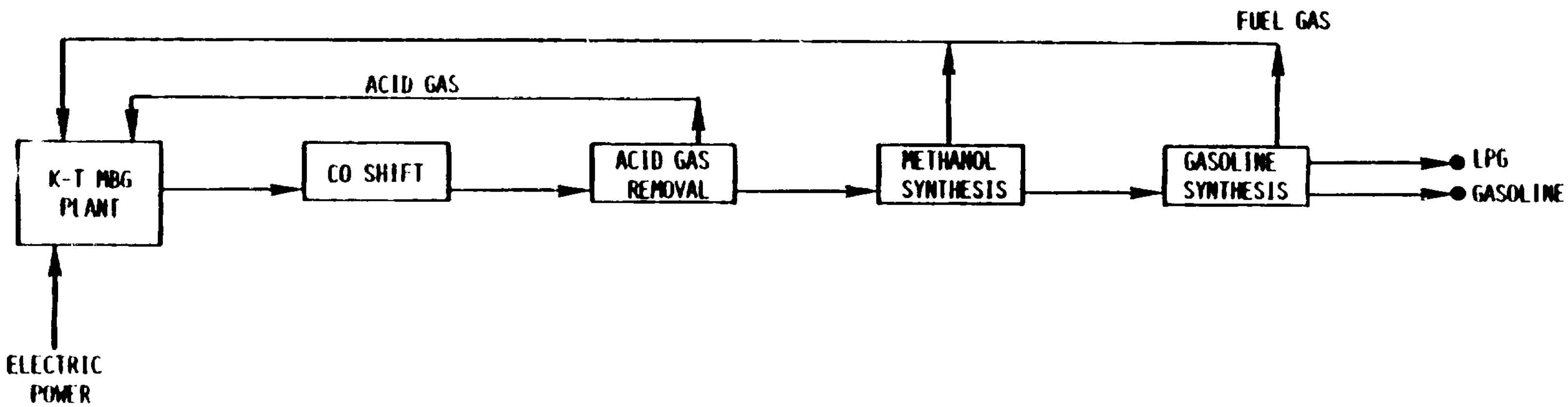


Figure VI.C.3. Koppers-Totzek to Gasoline

The Mobil M process has been selected over Fischer-Tropsch for a New Zealand synfuels complex. Based on this and a comparative economic study reviewed by the team, Mobil M was chosen for the gasoline synthesis process.

e. Hydrogen

The objective of this design is to produce high purity hydrogen (99%) for fuel-cells or hydrotreating service.

The required process steps include shift conversion to produce hydrogen, and acid gas removal to minimize CO_2 content.

The selected process is AGR followed by a high temperature shift and a pressure-swing-adsorption (PSA) process.

C. LURGI-METHANE AND LURGI-METHANE-METHANOL

1. Cost Evaluation

These cases were developed to examine the potential economic benefit of taking advantage of the high methane yield of the Lurgi gasifier by producing the methane as a product and converting the remaining gas to methanol. The cost results show that the mixed methane/methanol case results in a lower product cost per million BTU. The economic value of the two-product alternative depends on relative market prices for the two products, assuming there is a market for both. The product competitive evaluation in Chapter VIII indicates that methanol market prices may range from the same as methane, in direct competition for clean boiler fuel, to higher than methane as a substitute for distillate, or even higher as a gasoline blending stock. In the latter two cases, the combined methane/methanol plant would show a clear economic advantage over a methane only facility.

2. Description of Design

The Lurgi-to-methane has the same block flow diagram as the Koppers-Totzek to methane module described earlier. The main difference in performance is that the methane in the Lurgi medium BTU gas will pass unchanged through the shift, acid gas removal, methanation, and drying/compression stages. There is little methane in the Koppers-Totzek MBG.

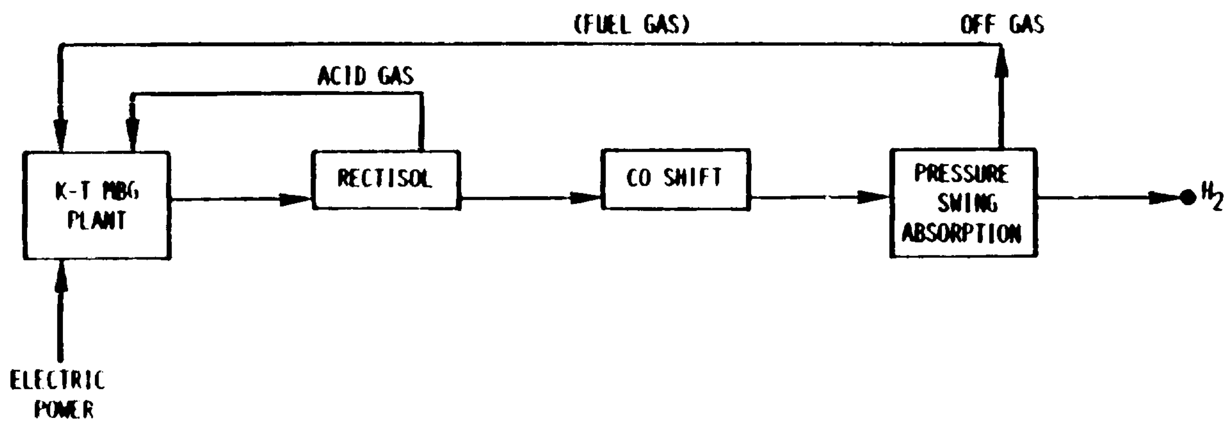


Figure VI.C.4. Koppers-Totzek to Hydrogen

	LURGI/METHANE	LURGI/METHANE & METHANOL
TOTAL CAPITAL INVESTMENT	3370	3532
OPERATIONS AND MAINTENANCE PLUS FEED	442	442
BY-PRODUCT REVENUE (COAL FINES)	47	47
90% YIELD (10^{12} BTU/year)	94.37	109.30
SNG	94.37	58.42
METHANOL	-0-	50.88
INTEGRATED PRICE ($1980/MMBTU)	$7.69	$6.81

NOTE: All costs in millions of dollars.

* Based on using the same price for methane in both cases.

Figure VI.C.5. Summary of Facility Costs for Lurgi Alternate Products

The block diagram for Lurgi to methane and methanol is shown in Figure VI.C.6. As before, the methane passes through unchanged. Acid gas removal is required for deep sulfur removal to avoid deactivating the methanol synthesis and methanation catalysts. As shown earlier, the methanol synthesis produces a "fuel gas" byproduct which is essentially excess hydrogen that occurs naturally in the Lurgi MBG. The CO_2 byproduct from acid gas removal is combined with this hydrogen in the methanation stage to yield methane and water. In the last stage, the water is removed and the gas is compressed to pipeline pressure.

D. KOPPERS-TOTZEK AND TEXACO COMBINED METHANE AND MBG

1. Introduction

The purpose of this analysis was to examine a facility that can produce either SNG or MBG, at the definitive design level. Two cases were examined; Koppers-Totzek gasification only, and mixed Koppers-Totzek and Texaco. The facility consists of four 5000 ton per day MBG modules feeding an upgrading plant producing MBG. In the first case, all four modules use Koppers-Totzek gasifiers. In the second case, the second, third, and fourth modules use Texaco gasifiers. The design guidelines are as follows:

- The first two facility modules must be designed to produce 100% MBG, 100% SNG, or a mixture of both.
- Any of the four facility modules must be capable of feeding the MBG Upgrading Plant.
- The MBG Upgrading Plant shall be integrated with the remainder of the Coal Gasification Facility, rather than being designed as an add-on plant.

2. Cost Evaluation

The cost figures show a clear economic advantage to incorporating the Texaco gasifiers. As described earlier, this results from the higher efficiency and lower operating cost of the Texaco gasifier, which

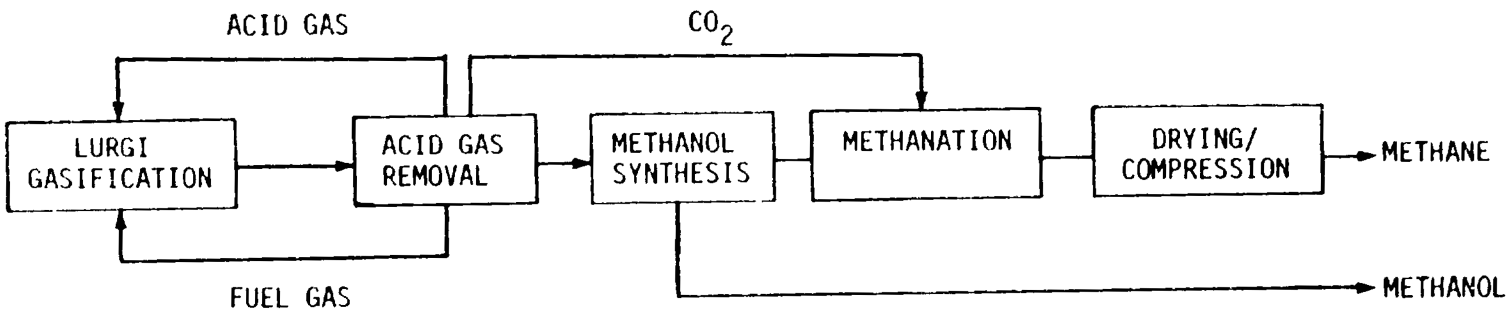


Figure VI.C.6. Lurgi to Methanol and Methane

	KOPPER-TOTZEK/SNG & MBG INTEGRATED FACILITY	K-T/TEXACO/SNG & MBG INTEGRATED FACILITY
INSTANT PLANT COSTS (IN MILLION DOLLARS)		
TOTAL FACILITY INVESTMENT	1735	1640
TOTAL CAPITAL REQUIREMENTS	2579	2407
O&M COSTS (IN MILLION DOLLARS)		
FEEDSTOCK & CATALYST & CHEMICALS	184	184
OTHER O&M	205	155
TOTAL O&M	389	339
MBG ANNUAL PRODUCT (10^{12} BTU)	45.04	51.45
SNG ANNUAL PRODUCT (10^{12} BTU)	35.04	37.62
INTEGRATED FACILITY PRODUCT PRICE ($/MMBTU) (IN CONSTANT 1980 DOLLARS)	$ 8.02	$ 6.49

NOTES:

(1) ALL O&M COSTS, ANNUAL PRODUCT AND PRODUCT PRICES ARE FOR THE FACILITY AT 90 PERCENT OPERATING CAPACITY.

(2) ALL COSTS ARE IN MILLIONS OF DOLLARS UNLESS OTHERWISE NOTED.

Figure VI.D.1. Summary of Facility Costs for Integrated Candidate Processes

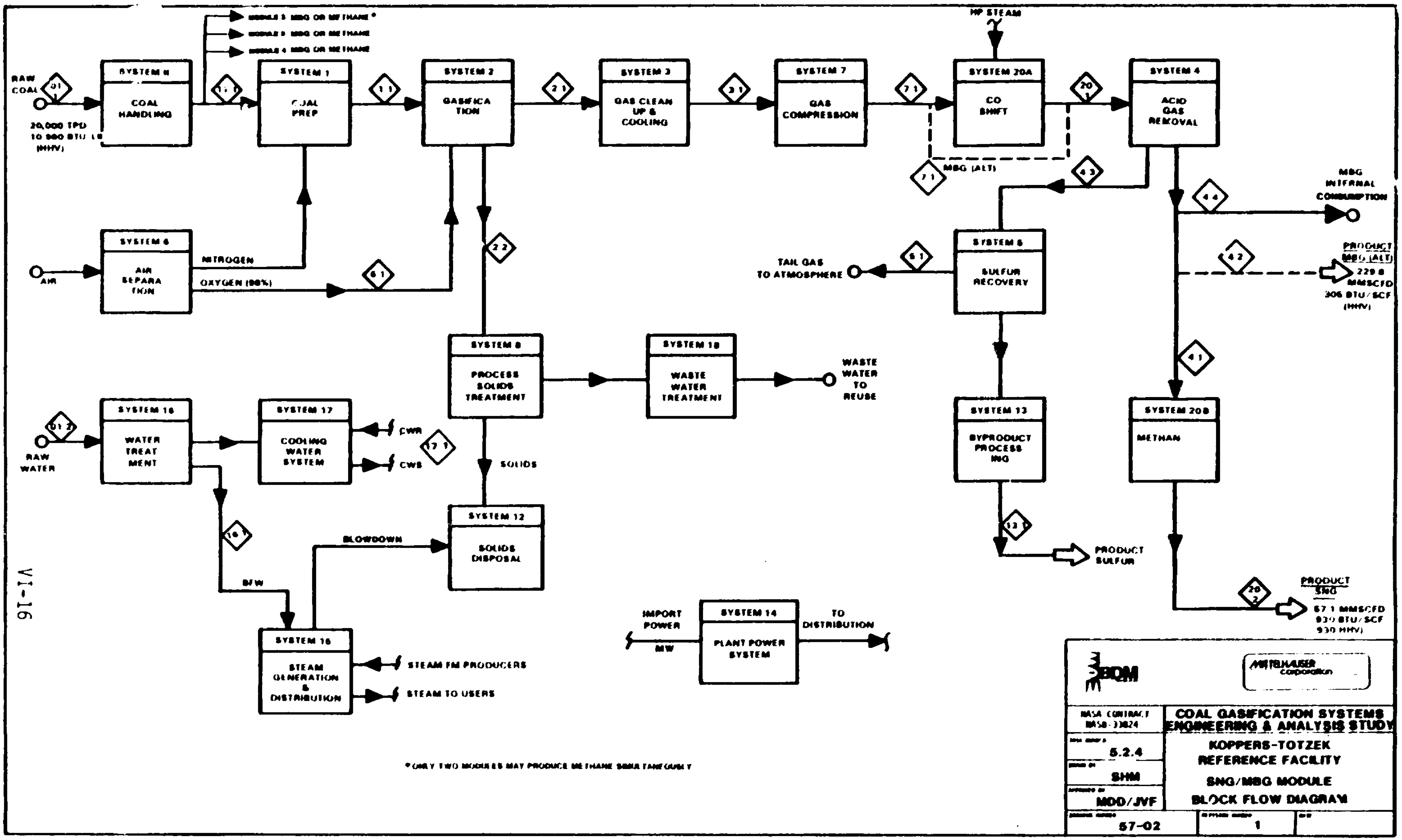
more than compensates for its higher capital cost. The fraction of annual BTU's going to MBG or methane is slightly different in the two cases, reflecting small differences in gas composition and gas stream conditions.

There is a capital cost "penalty" associated with the desired flexibility to use any of the four modules with the upgrading plant and to make up to 100% SNG. As described below, the Acid Gas Removal Systems in all four modules are specified to achieve deep sulfur removal (to avoid damaging catalysts in the upgrading units), although only two modules would supply MBG for upgrading at any one time.

3. Description of Design

The resulting integrated Koppers-Totzek and Texaco modules are illustrated in Figure VI.D.2. and VI.D.3. The sequence of processing steps to upgrade MBG to methane is shift, followed by acid gas removal, followed by methanation. For Koppers-Totzek the gas is compressed prior to the shift. The designs and design tradeoffs are described in detail in Appendix G.

The design task began with the already-completed Reference Facility Designs, described in Chapter III. These designs were examined for integration possibilities in the light of other previous analyses of SNG as an alternate product from K-T and Texaco gasification. The examination revealed that the major candidate for system integration was System 4, Acid Gas Removal. The manufacture of SNG requires a methanation catalyst which has little or no sulfur tolerance, therefore essentially complete removal of all sulfur compounds is required upstream of the methanation system. For this reason, the Acid Gas Removal Systems in the Reference Facility Designs, which had been designed for a 200 ppmv treated gas sulfur concentration, were replaced by systems designed for a maximum treated gas sulfur concentration of 1 ppmv. The remainder of the sulfur removal was designed to be done in the Methanation System. Because of the requirement that each module be capable of feeding the MBG Upgrading Plant, all of the Acid Gas Removal System in each facility



RAW COAL
20,000 TPD
(HHV)
SYSTEM 6
COAL HANDLING
SYSTEM 1
COAL PREP
SYSTEM 2
GASIFICATION
SYSTEM 3
GAS CLEAN UP & COOLING
SYSTEM 7
GAS COMPRESSION
HP STEAM
SYSTEM 20A
CO SHIFT
SYSTEM 4
ACID GAS REMOVAL
MBG (ALT)
MBG INTERNAL CONSUMPTION
SYSTEM 6
AIR SEPARATION
AIR
NITROGEN
OXYGEN (98%)
TAIL GAS TO ATMOSPHERE
SYSTEM 5
SULFUR RECOVERY
PRODUCT MBG (ALT)
306 BTU/SCF (HHV)
SYSTEM 8
PROCESS SOLIDS TREATMENT
SYSTEM 18
WASTE WATER TREATMENT
WASTE WATER TO REUSE
SYSTEM 13
BYPRODUCT PROCESSING
SYSTEM 20B
METHAN
SYSTEM 16
WATER TREATMENT
RAW WATER
SYSTEM 17
COOLING WATER SYSTEM
CWR
CWS
SOLIDS
SYSTEM 12
SOLIDS DISPOSAL
BLOWDOWN
BFW
PRODUCT SULFUR
PRODUCT SNG
IMPORT POWER
MW
SYSTEM 14
PLANT POWER SYSTEM
TO DISTRIBUTION
SYSTEM 15
STEAM GENERATION & DISTRIBUTION
STEAM FM PRODUCERS
STEAM TO USERS
BDM
NASA CONTRACT NAS8-33824
5.2.4
SHM
MDD/JVF
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
KOPPERS-TOTZEK REFERENCE FACILITY
SNG/MBG MODULE
BLOCK FLOW DIAGRAM
57-02
1

Figure VI-D-2

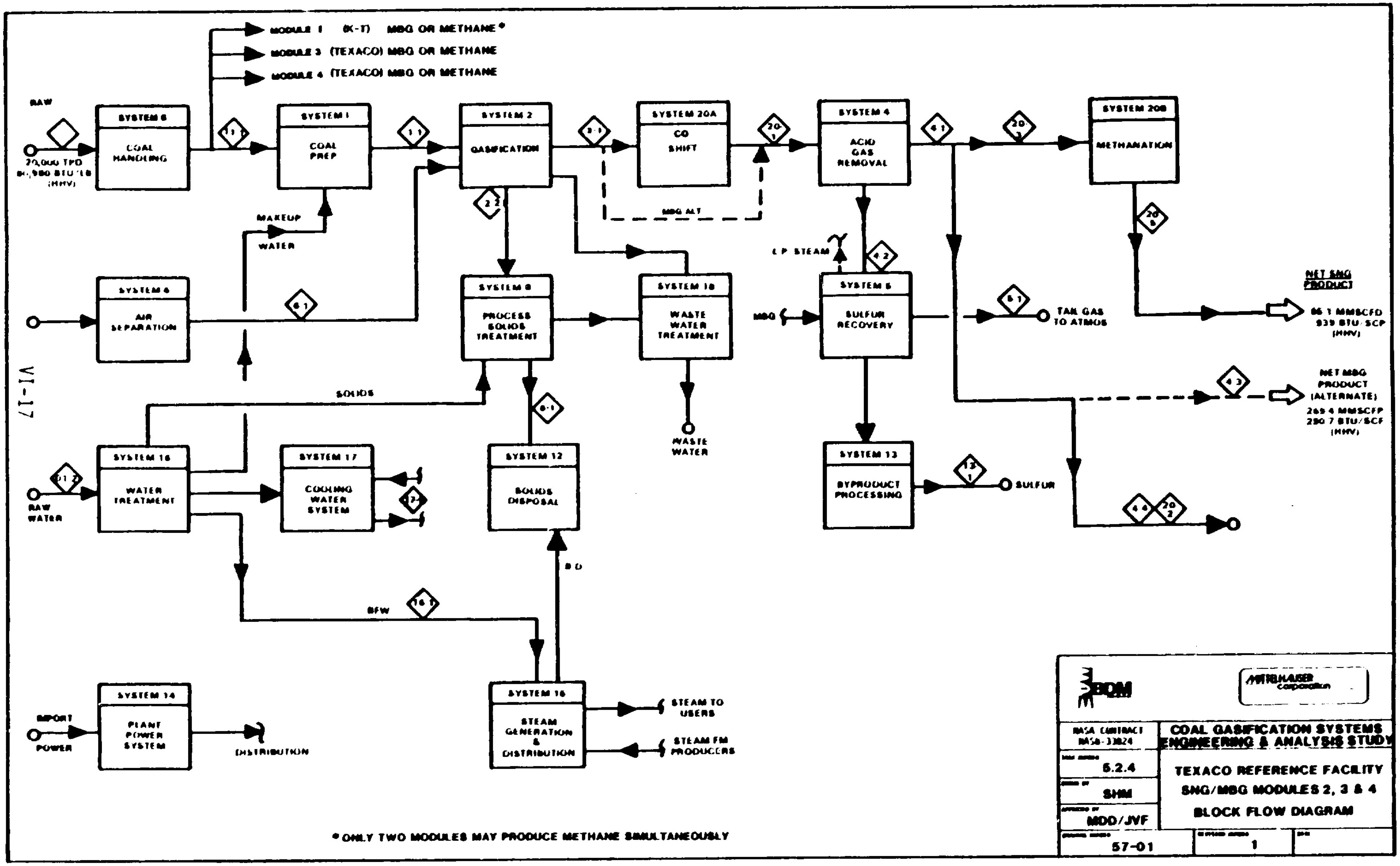

MODULE 1 (K-T) MBG OR METHANE*
MODULE 3 (TEXACO) MBG OR METHANE
MODULE 4 (TEXACO) MBG OR METHANE
RAW
20,000 TPD
BTU/LB
(HHV)
SYSTEM 5
COAL HANDLING
SYSTEM 1
COAL PREP
SYSTEM 2
GASIFICATION
SYSTEM 20A
CO SHIFT
MBG ALT
SYSTEM 4
ACID GAS REMOVAL
SYSTEM 20B
METHANATION
MAKEUP
WATER
SYSTEM 6
AIR SEPARATION
SYSTEM 8
PROCESS SOLIDS TREATMENT
SYSTEM 18
WASTE WATER TREATMENT
L P STEAM
MBG
SYSTEM 5
SULFUR RECOVERY
TAIL GAS TO ATMOS
NET SNG PRODUCT
939 BTU/SCF
(HHV)
NET MBG PRODUCT (ALTERNATE)
(HHV)
SOLIDS
WASTE WATER
SYSTEM 16
WATER TREATMENT
RAW WATER
SYSTEM 17
COOLING WATER SYSTEM
SYSTEM 12
SOLIDS DISPOSAL
SYSTEM 13
BYPRODUCT PROCESSING
SULFUR
BD
BFW
SYSTEM 14
PLANT POWER SYSTEM
IMPORT POWER
DISTRIBUTION
SYSTEM 15
STEAM GENERATION & DISTRIBUTION
STEAM TO USERS
STEAM FM PRODUCERS
BDM
NASA CONTRACT
5.2.4
SHM
MDD/JVF
57-01
1
COAL GASIFICATION SYSTEMS ENGINEERING & ANALYSIS STUDY
TEXACO REFERENCE FACILITY
SNG/MBG MODULES 2, 3 & 4
BLOCK FLOW DIAGRAM
* ONLY TWO MODULES MAY PRODUCE METHANE SIMULTANEOUSLY

Figure VI-D-3

were replaced. Integration of other systems between the MBG Facility and the Upgrading Plant was purposely minimized to allow for the MBG/SNG flexibility desired by TVA.

CHAPTER VII
SCHEDULING ANALYSIS

A. REFERENCE FACILITY SCHEDULES

The TVA gasification plant project is designed to be completed in increments based on a four independent module concept. While some systems serve the entire facility, the design is such that each module can be operated alone. Also, Module 1 will be completed and operating prior to completion of Module 2 and so forth for the total facility.

A system level schedule assessment has been completed. Figures VII-A-1 through VII-A-7 present the result of this analysis. It should be noted that this analysis is based on general experience being applied to the subject project. There has been no definitive investigation into such items as local labor constraints, current shop lead times and other factors likely to influence the final project schedule. Nevertheless, these schedules are appropriate for early project planning and can serve to highlight potential bottlenecks or constraints. Dotted extensions of activity schedule estimates serve to illustrate the effects of longer completion times.

In addition to presenting activity time requirements, the schedules also illustrate intersystem dependence. Items which are assessed to be critical with regard to schedule completion have been emphasized by heavy lines. Overall, the project schedule is based on selecting an A/E contractor in the fourth quarter of 1980. Completion of construction and testing of Module 4 is then scheduled for mid-1987.

The following discussion elaborates on the results of the analysis and the interpretation of the figures.

1. Milestones

The program development methodology encompasses the establishment of a specific set of time structured elements scheduled for completion at predesignated dates. To facilitate effective program management of system development, and to ensure management review of program status, a set of

objective oriented milestones have been established. These milestones include:

(1) Program Requirements Review (PRR)
(2) Preliminary Design Review (PDR)
(3) Critical Design Review (CDR)
(4) Operational Readiness Review (ORR)
(5) Start of Commercial Operations (SCO)

a. Program Requirements Review (PRR)

The PRR will be a vehicle for review and approval of the complete systems requirements for all functions to be performed by the coal gasification facility. It will occur four months from the start date and will present for program management approval a complete Functional Description, a Test Plan and a list of system deliverables related to both the total systems and individual module development.

b. Preliminary Design Review (PDR)

The PDR will occur twelve months after the start date and at this time program management will review the complete system and subsystem designs. All system and subsystem specifications will be completed in draft form for review. The Test Requirements will be approved at this review. Construction of well defined systems such as coal handling, solids disposal, plant power, general facilities may begin shortly after the PDR and prior to the critical design review.

c. Critical Design Review (CDR)

Twenty months from the start date, a CDR will be held to approve all specifications. The drafts presented at the PDR will be revised as necessary to meet program development requirements, and specifications will be defined to the subsystem level. The final version of system, subsystem, and test specifications will be approved at the CDR. Approval of the CDR will mark the initiation of major construction activity for all systems not already started. The final designs and specifications provide the necessary guidance and instructions for remaining program development activities.

d. Operational Readiness Review (ORR)

This milestone is the fourth to be reached and occurs approximately 51 months from the program start date. The objective of the ORR is to review completed system acceptance test results to determine operational readiness of each module. Complete program documentation review is also performed during this review. Following the ORR a six month period of module testing will commence.

e. Start of Commercial Operation (SCO)

The SCO constitutes the final phase of program development. The results of module testing and evaluation will be reviewed and commercial operation of each module will commence. Total facility management, operation, maintenance, and logistic support will proceed in accordance with the conceptualized standard operating procedures, facility operating instruction, system safety plans, and quality assurance requirements.

2. Master Schedule

The major program development activities and their time phased relationship to each of the four system modules is shown in the Coal Gasification Facility Project Master Schedule (Figure VII-A-2). Specific major activities include engineering procurement, construction, and testing. Also included are the program milestones and their associated dates.

B. LOGIC NETS

The following schedule logic nets have been prepared:

(1) Summary Diagram. This shows project milestones and an overview of the engineering, procurement, construction and testing of the total facility.

(2) Module I General Facilities and Offsite Systems. Engineering, procurement, construction and test phases are shown.

(3) Modules I-IV. Individual nets are given for the engineering through test cycle.

(4) Management and Planning Functions. This net addresses contract monitoring selection of AIEs, and other management and planning functions.

(5) Design, Procurement, Construction. The major activities in designing, procuring and constructing the facility are scheduled.

There are two major concerns with regard to the overall plan:

(1) It is important to insure that all possible early construction is completed before the last equipment arrives, i.e., maintain overlap between delivery and construction.

(2) Gasification, gas cleaning, and acid gas removal are time-consuming to test, and will require a relatively long time before attaining design scale equilibrium.

Other critical schedule factors were identified:

(1) Module I - General facilities and offsite systems. There is a need for systems testing for cost handling, solids disposal and byproduct processing beyond what is shown. This testing will have to be proportionately more than for the plant power system.

(2) Module I - Engineer/procure/construct/test. Gasification is the most critical function, particularly when needed testing is added.

(3) Module II, III and IV. The Gas Cleanup/Cooling system is the most critical. The systems testing requirement may not afford time for slippage or adequate testing. Procurement and construction might be started earlier to ease the tight schedule.

TVA COAL GASIFICATION MBG

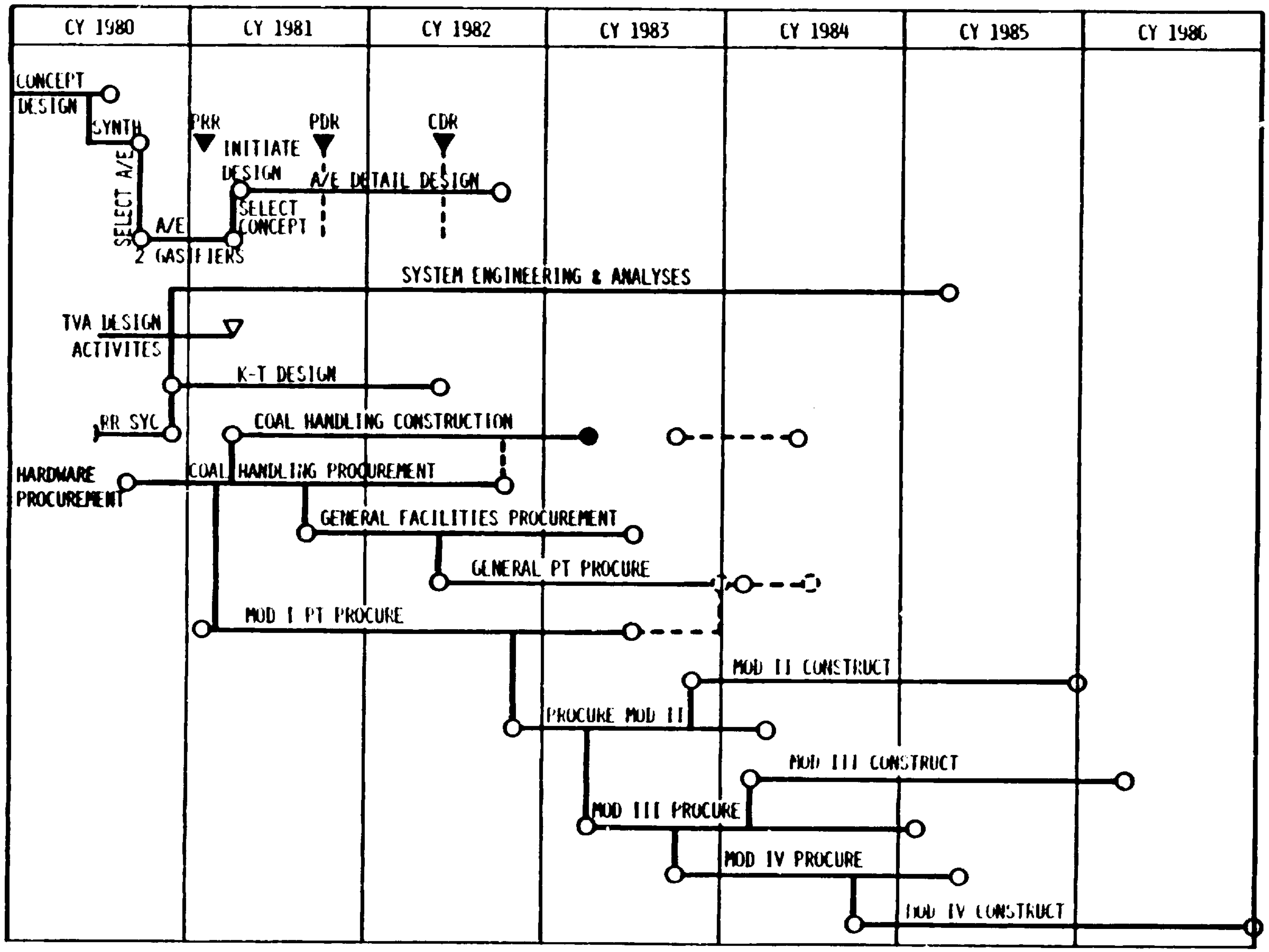
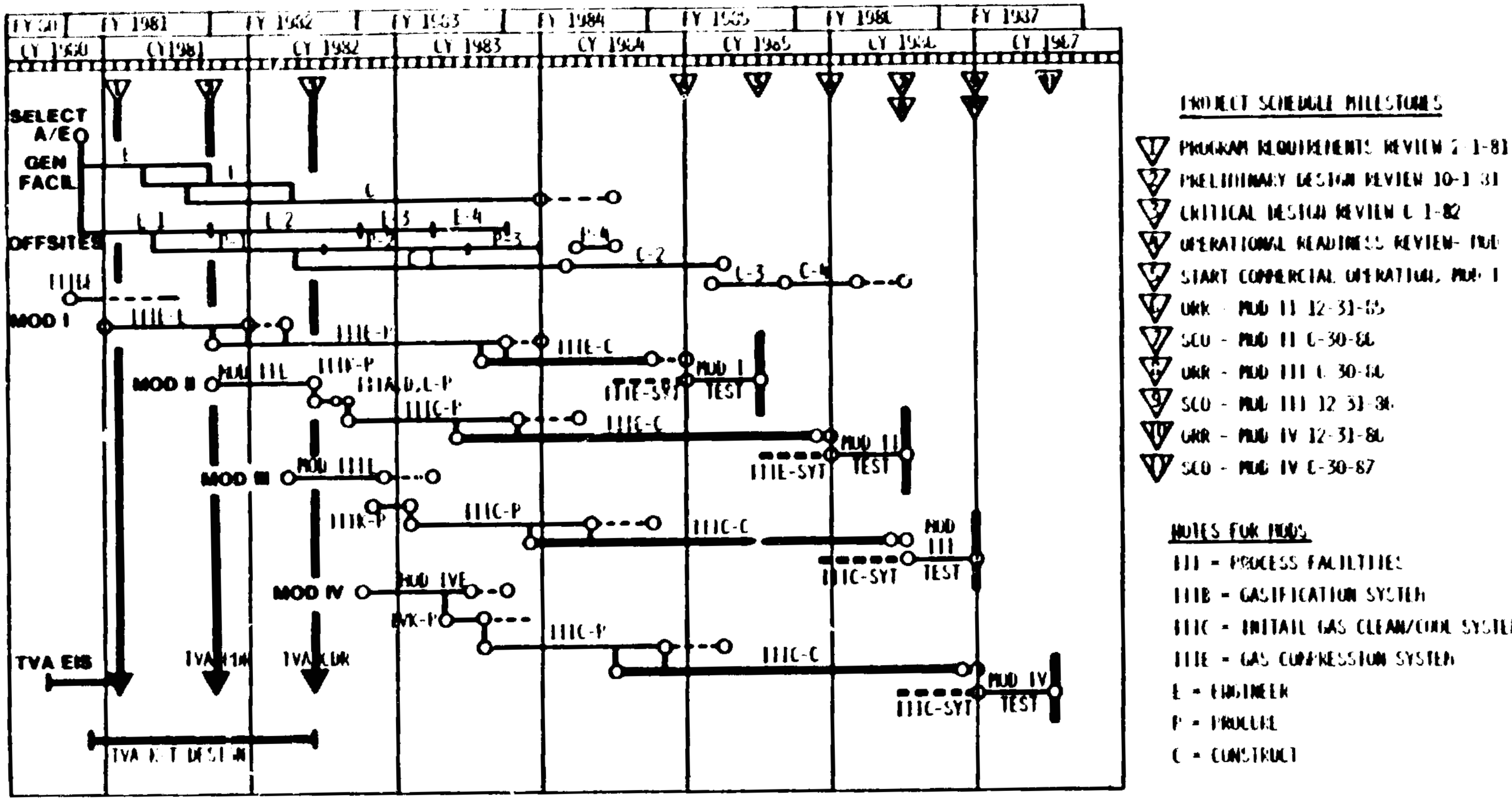


Figure VII-A-1. Overall Project Schedule

TVA SCHEDULING FOR MBG PLANT SUMMARY DIAGRAM



CRITICAL ITEMS ONLY:

SEE MODULE SHEETS FOR ENTIRE LISTING, INCLUDING COMMENTS. TWO INHERENT THINGS TO WATCH FOR:

(1) INSURE THAT ALL POSSIBLE EARLY CONSTRUCTION IS COMPLETED BEFORE LAST EQUIPMENT ARRIVES, I.E., MAINTAIN OVERLAP.

(2) ITEMS B (GASIFICATION), C (GAS CLEANUP), D (ACID GAS REMOVAL) ARE TIME-CONSUMING TO TEST, WILL REQUIRE RELATIVELY LONG TIME BEFORE ATTAINING DESIGN-SCALE EQUILIBRIUM.

Figure VII-A-2. Master Schedule

MODULE I SCHEDULE WITH RELATED GENERAL FACILITIES ENGINEERING (E), PROCUREMENT (P) AND CONSTRUCTION (C)

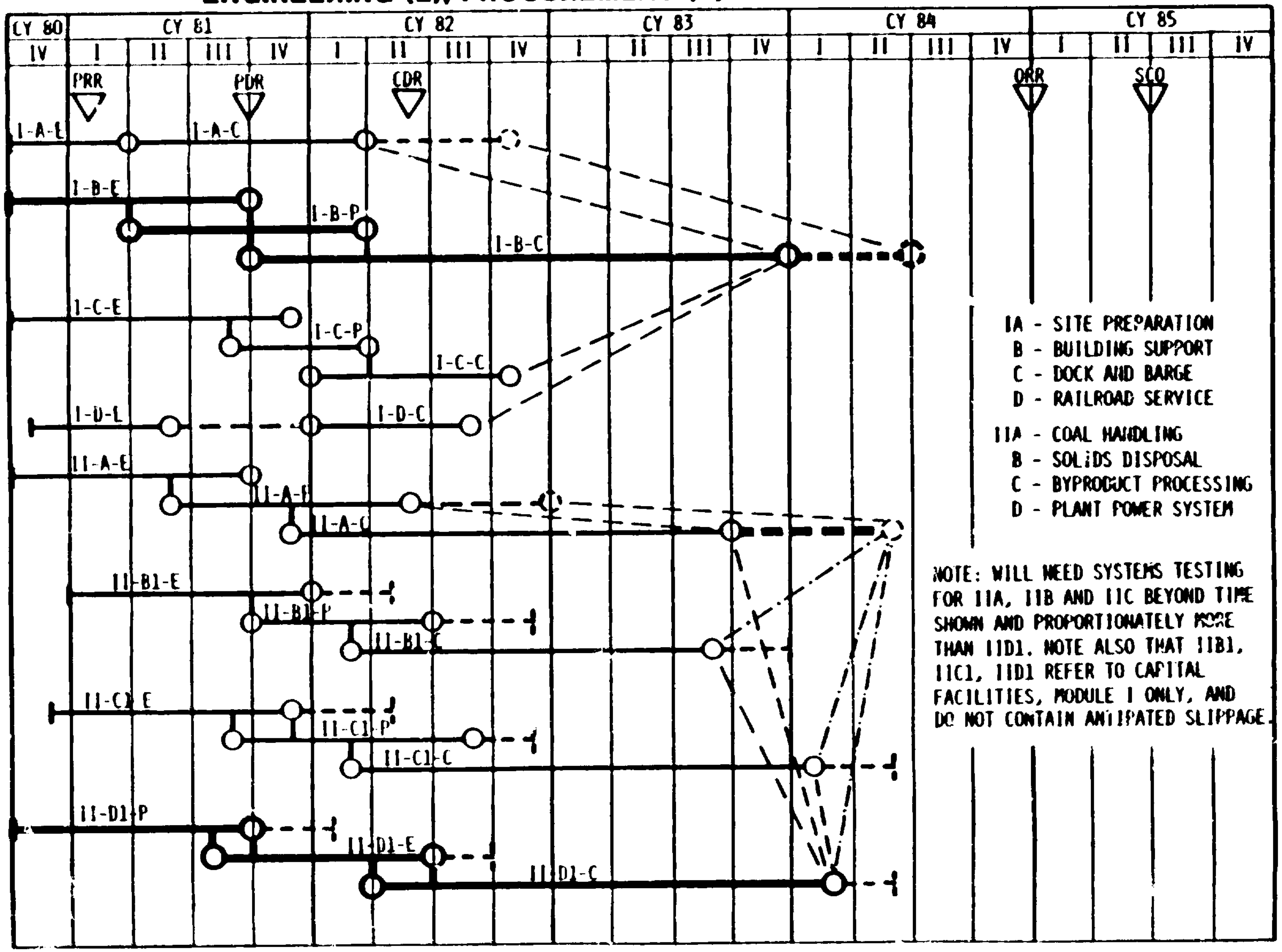


Figure VII-A-3. General Facilities Schedule

MODULE I ENGINEER/PROCURE/CONSTRUCT/TEST

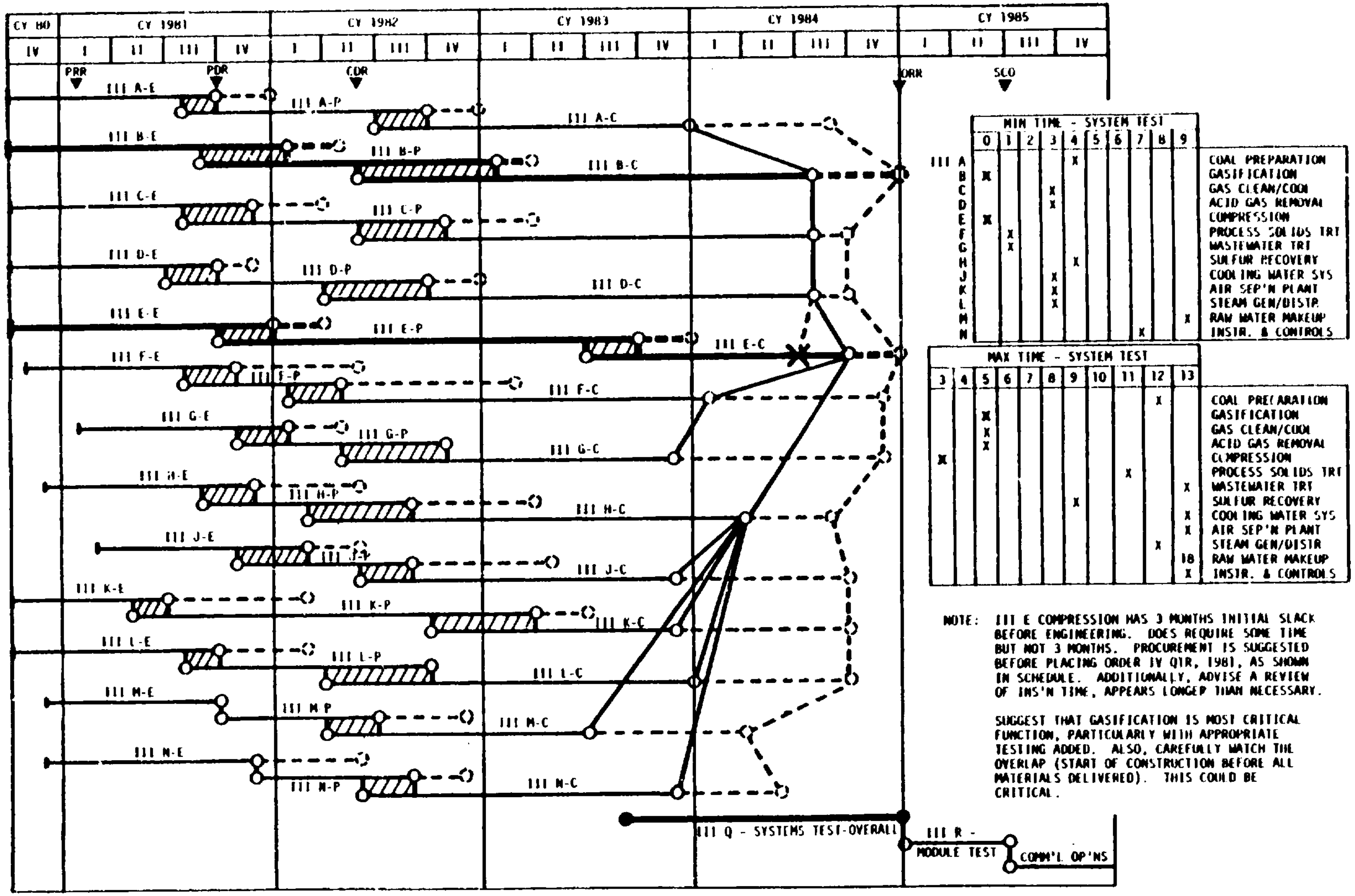
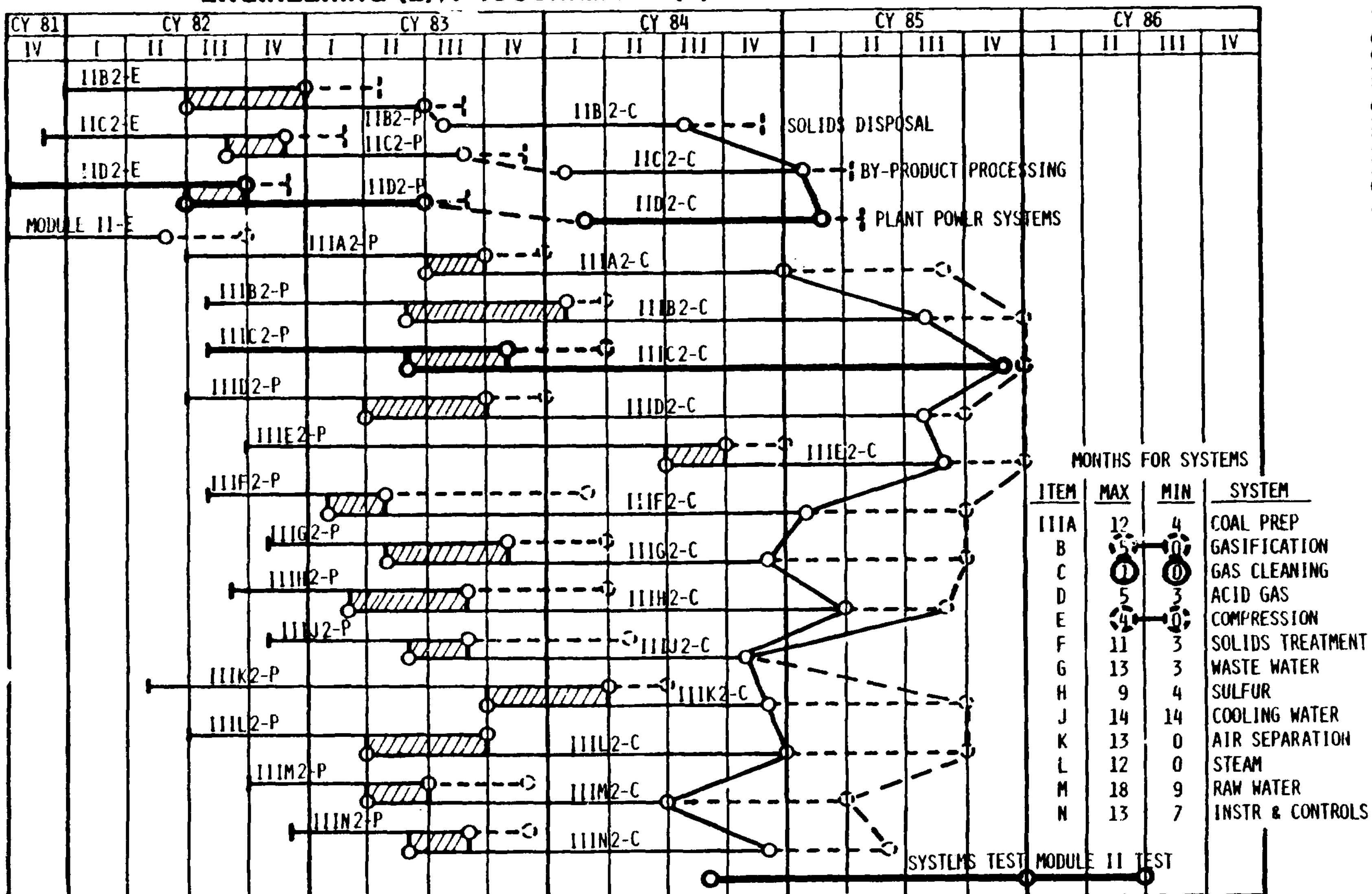


Figure VII-A-4. Module I Schedule

MODULE II SCHEDULE WITH RELATED GENERAL FACILITIES
ENGINEERING (E), PROCUREMENT (P) AND CONSTRUCTION (C)



MONTHS FOR SYSTEMS

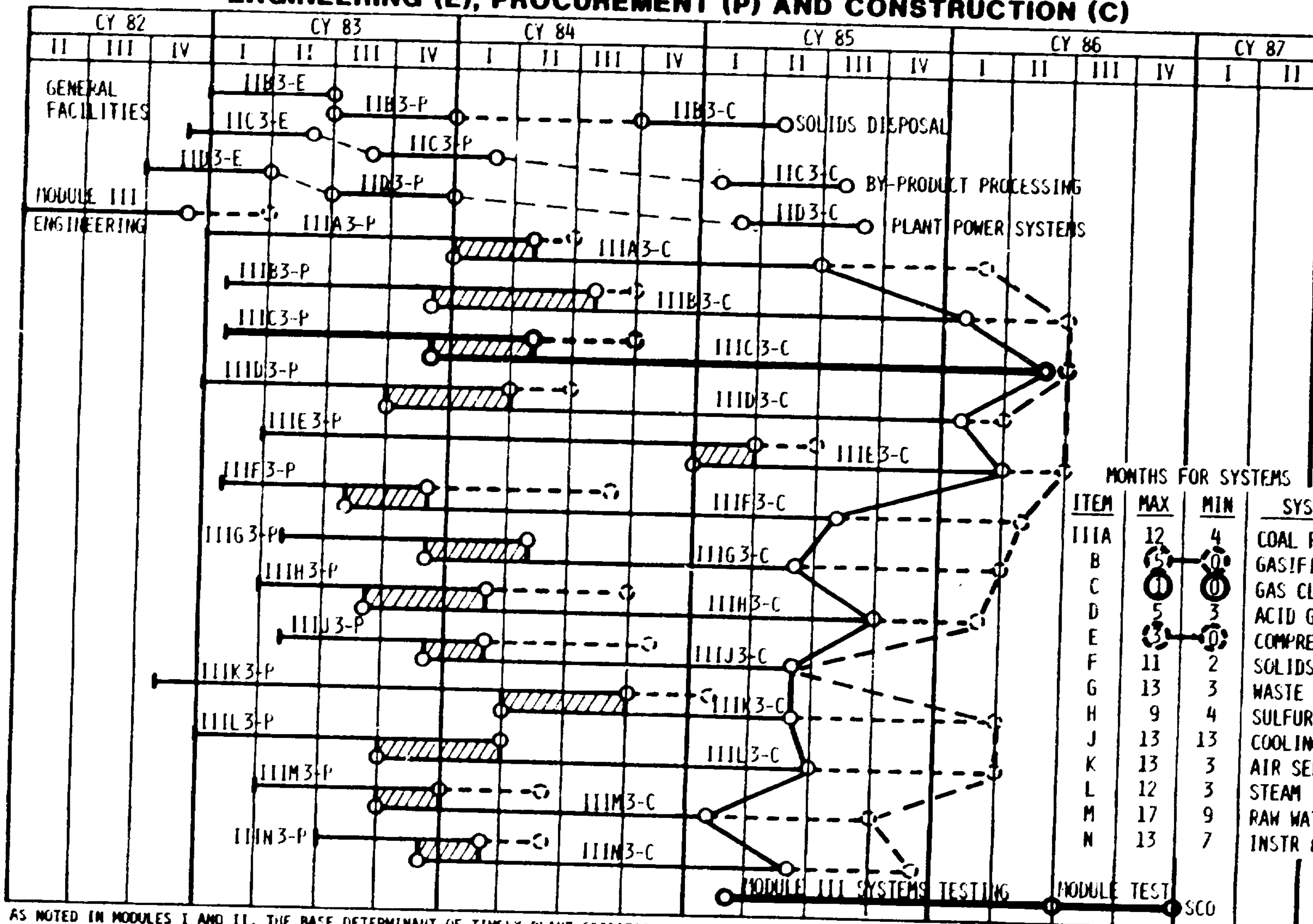
ITEM	MAX	MIN	SYSTEM
IIIA	12	4	COAL PREP
B	5	0	GASIFICATION
C	1	0	GAS CLEANING
D	5	3	ACID GAS
E	4	0	COMPRESSION
F	11	3	SOLIDS TREATMENT
G	13	3	WASTE WATER
H	9	4	SULFUR
J	14	14	COOLING WATER
K	13	0	AIR SEPARATION
L	12	0	STEAM
M	18	9	RAW WATER
N	13	7	INSTR & CONTROLS

NOTE: REQUIREMENT FOR SYSTEMS TESTING OF GAS CLEANUP/COOLING MAY NOT AFFORD TIME FOR ANY SLIPPAGE OR FOR ADEQUATE TESTING. HOWEVER, PROCUREMENT AND CONSTRUCTION MAY BE INITIATED EARLIER IN PLANNED CONSTRUCTION PERIOD OF MODULE II PROCESSING FACILITILS

IN GENERAL FACILTIES IID - PLANT POWER SYSTEMS - 8 MONTH HIATUS BETWEEN PURCHASE COMPLETION AND CONSTRUCTION START DOES NOT APPEAR NECESSARY.

Figure VII-A-5. Module II Schedule

MODULE III SCHEDULE WITH RELATED GENERAL FACILITIES ENGINEERING (E), PROCUREMENT (P) AND CONSTRUCTION (C)



MONTHS FOR SYSTEMS

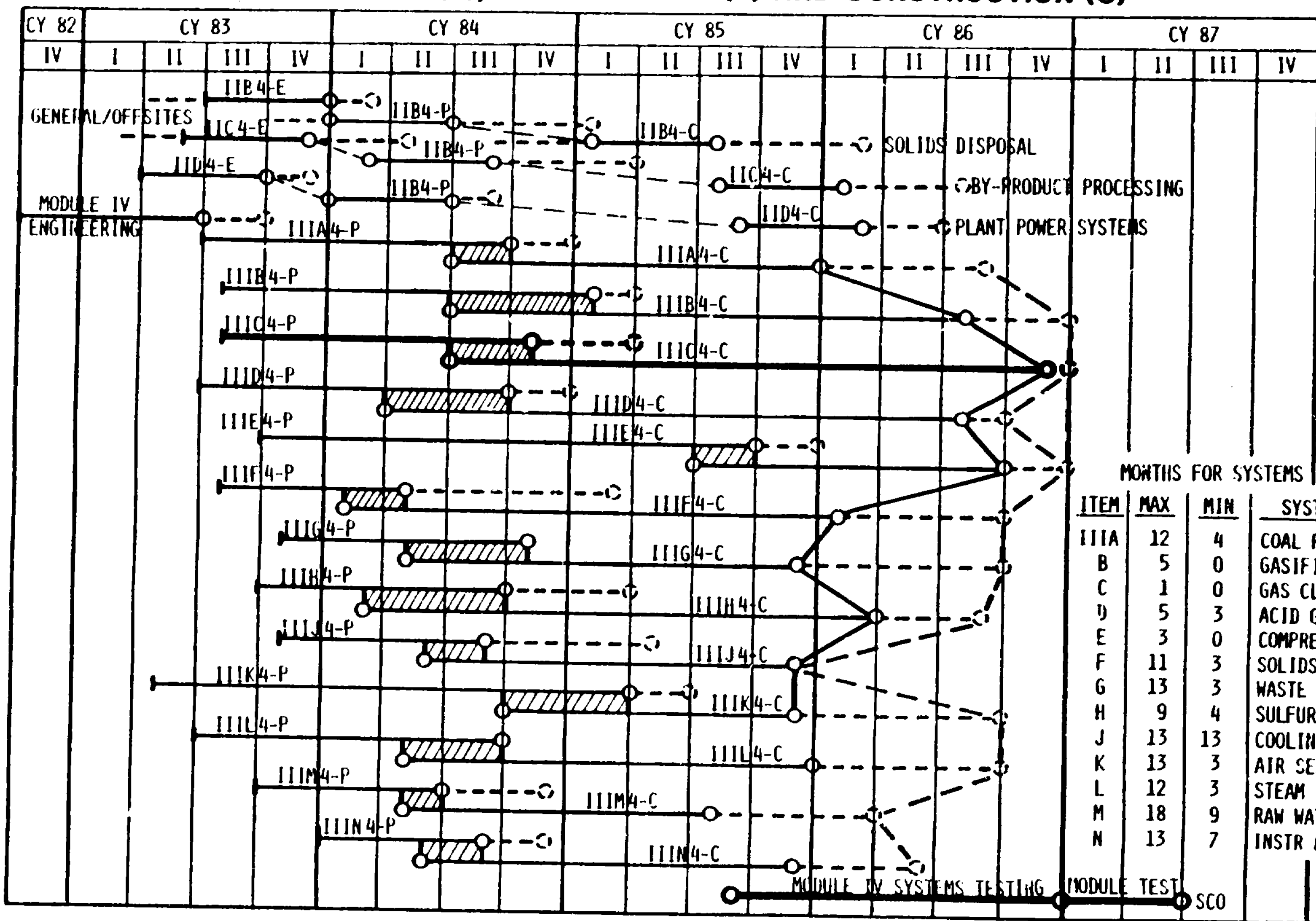
ITEM	MAX	MIN	SYSTEM
IIIA	12	4	COAL PREP
B	5	0	GASIFICATION
C	0	0	GAS CLEANING
D	5	3	ACID GAS
E	3	0	COMPRESSION
F	11	2	SOLIDS TREATMENT
G	13	3	WASTE WATER
H	9	4	SULFUR
J	13	13	COOLING WATER
K	13	3	AIR SEPARATION
L	12	3	STEAM
M	17	9	RAW WATER
N	13	7	INSTR & CONTROLS

AS NOTED IN MODULES I AND II, THE BASE DETERMINANT OF TIMELY PLANT OPERATIONS WILL BE SUFFICIENT TIME AVAILABLE FOR TESTING, EVALUATING AND DEBUGGING PROCESS FACILITIES. INITIAL GAS CLEANUP AND COOLING STILL APPEARS TO BE MOST SENSITIVE FUNCTIONAL AREA WITH COMPRESSION AND GASIFICATION NEXT MOST CRITICAL. IT IS BELIEVED THAT INSTALLATION TIME SHOWN FOR COMPRESSION IS MORE THAN ADEQUATE WITH APPROPRIATE SCHEDULING, TO RECEIVE, INSTALL, CONNECT, TEST AND INTEGRATE INTO MODULE PROCESS FLOW.

ALSO AS NOTED BEFORE, THE APPARENT LULLS BETWEEN ENGINEERING, PROCUREMENT AND CONSTRUCTION FOR THE RELATED GENERAL FACILITIES DOES NOT APPEAR REALISTIC, HOWEVER, IT DOES RAISE A QUESTION ABOUT FRONT-LOADING THESE ACTIVITIES TO SAVE CONSTRUCITON TIME AND INFLATIONARY ESCALATIONS.

Figure VII-A-6. Module III Schedule

MODULE IV SCHEDULE WITH RELATED GENERAL FACILITIES ENGINEERING (E), PROCUREMENT (P) AND CONSTRUCTION (C)



MONTHS FOR SYSTEMS

ITEM	MAX	MIN	SYSTEM
IIIA	12	4	COAL PREP
B	5	0	GASIFICATION
C	1	0	GAS CLEANING
D	5	3	ACID GAS
E	3	0	COMPRESSION
F	11	3	SOLIDS TRTMNT
G	13	3	WASTE WATER
H	9	4	SULFUR
J	13	13	COOLING WATER
K	13	3	AIR SEPAR'N
L	12	3	STEAM
M	18	9	RAW WATER
N	13	7	INSTR & CONTR

THE SAME ELEMENTS APPLY TO MODULE IV AS FOR PRIOR FACILITIES.

Figure VII-A-7. Module IV Schedule

CHAPTER VIII
PRODUCT COMPETITIVE EVALUATIONS

A. Introduction and Background

The purpose of this task is to provide a preliminary assessment of the potential competitiveness of the candidate products of the TVA coal gasification plant. The analysis is based on projected national average prices for competing fuels, and comparisons of these prices with projected product costs for the gasification plant. This analysis does not address the potential size of the market. Additionally, transportation and distribution costs of the gasification products are not included in the comparisons.

B. Estimated Gasification Plant Product Prices

The estimated product prices for gasification plant products are shown in Figure VIII.B.1. These prices are developed in the designs and cost estimates documented in Chapter III, V, and VI. The prices are expressed in 1980 dollars and represent the price in constant 1980 dollars that would recover the cost of service of the plant. Thus, the corresponding nominal or current price would increase in proportion to the general rate of inflation.

C. Selection of Competing Fuels

Figure VIII.C.1. summarizes the rationale for the selection of fuels with which the gasification plant products might compete. Medium BTU gas (MBG) would compete with other industrial boiler fuels. Methane would compete with other sources of new gas supplies for gas utilities. The highest price a gas utility would pay for new gas would be determined in part by the highest priced competing fuel. Distillate for space heating is by far the most significant highly priced fuel competing with natural gas.

Methanol has a wide variety of uses. It can compete with distillate and natural gas as a boiler fuel, turbine fuel and chemical feedstock. Additionally, it can be blended into gasoline or used as a pure motor fuel. Use of methanol for all these applications is expected to grow dramatically over the next ten years. Methanol can also be converted to gasoline.

GASIFIER	MBG PRICE
K-T	6.64
TEXACO	5.00
BABCOCK & WILCOX	6.39
SLAGGING LURGI	4.31
LURGI	5.44

ALTERNATE PRODUCTS	KOPPERS-TOTZEK	TEXACO
METHANE	8.03	7.63
METHANOL	8.08	7.54
GASOLINE	11.21	9.04

Figure VIII.B.1. Gasification Plant Product Costs, $1980/MMBTU

GASIFICATION PLANT PRODUCT	COMPETING FUEL	RATIONALE
MBG	INDUSTRIAL RESIDUAL FUEL OIL	COMPETE FOR BOILER FUEL
	INDUSTRIAL DISTILLATE	COMPETE FOR BOILER FUEL
METHANE	NEW NATURAL GAS	INCREMENTAL COST OF GAS SUPPLY
	RESIDENTIAL DISTILLATE	COMPETE FOR SPACE HEATING
	COMMERCIAL DISTILLATE	COMPETE FOR SPACE HEATING
METHANOL	INDUSTRIAL DISTILLATE	COMPETE FOR BOILER AND TURBINE FUEL
	NEW NATURAL GAS	COMPETE FOR BOILER FUEL AND CHEMICAL FEED
	WHOLESALE GASOLINE	COMPETE FOR GASOHOL BLENDING STOCK OR MOTOR FUEL
GASOLINE	WHOLESALE GASOLINE	

Figure VIII.C.1. Selection of Competing Fuels

D. Projected Prices for Competing Fuels

Prices for competing fuels are tabulated in Figure VIII.D.1, 2, and 3. Projected world oil prices are listed in Figure VIII.D.7. Except where otherwise specified, prices are taken from the Energy Information Administration 1979 Annual Report to Congress, Vol. 3, DOE/EIA-0173(79)13. All oil and gas prices are expressed in 1980 dollars per million btu. Power is given in 1980 cents per kilowatt hour. In most cases, the low, medium and high prices listed by EIA for each year are shown (these do not correspond to specific cases given by EIA, but represent a composite).

World oil prices range from no real increase in the low scenario to a doubling of the real price in the high scenario over the operating life of the plant. Fuel oil, distillate and gasoline show similar ranges in Figure VIII.D.8. Power cost variations and real growth are very low due to the high portion of costs represented by capital recovery and to the large existing capital base relative to projected growth. The "wholesale gasoline" prices are taken as 90% of retail, based on recent EIA data showing wholesale gasoline at 88% to 91% of retail.

New natural gas real price increases are projected to range from 25% to over 300% over the operating life of the plant, as shown in Figure VIII.D.1. Average residential, industrial and commercial gas prices and distillate prices for space heating are shown in Figure VIII.D.1. The various gas prices are similar to one another, as are the residential and commercial distillate.

E. Comparison of Gasification Plant Product Prices with Competing Products

The prices of gasification plant products are compared with high and low projected prices for competing fuels in Figures VIII.D.3, D.4, D.5 and D.6. All prices are in 1980 dollars. World crude price is also displayed in Figure VIII.D.3 for reference. For the sale of clarity, only the Koppers-Totzek gasification plant product prices are displayed. Other prices from Figure VIII.B.1. are readily compared, however, since the 1980 dollar prices for gasification plant products

	1985			1990			1995		
	LO	MID	HI	LO	MID	HI	LO	MID	HI
NEW INTERSTATE	2.97	2.97	2.97	3.70	3.72	4.00	4.18	4.21	4.37
NEW INTRASTATE	3.10	3.10	3.10	3.90	3.91	4.20	4.39	4.39	4.60
CANADA	4.71	5.67	6.99	4.71	6.62	7.44	4.71	7.37	9.57
MEXICO	3.85	4.56	5.56	3.85	5.27	6.27	3.73	4.61	5.74
LNG	4.30	4.83	5.29	3.91	4.67	5.28	3.73	4.61	5.74
MIN/MAX	2.97		6.99	3.70		7.44	3.73		9.57

Figure VIII.D.1. Projected New Natural Gas Prices, $1980/MMBTU

	1. RESIDENTIAL GAS			2. INDUSTRIAL GAS			3. COMMERCIAL GAS		
	LO	MID	HI	LO	MID	HI	LO	MID	HI
1985	3.81	3.83	3.86	3.36	3.47	3.56	3.31	3.33	3.37
1990	4.40	4.65	4.86	4.06	4.85	4.90	3.93	4.19	4.37
1995	4.68	5.06	5.47	4.42	5.40	5.78	4.21	4.60	4.97
2000	4.46	4.67	4.72	4.75	4.96	5.01	4.50	4.71	4.76
2010	4.92	5.22	5.23	5.21	5.51	5.52	4.96	5.26	5.27

	4. RESIDENTIAL DISTILLATE			5. COMMERCIAL DISTILLATE		
	LO	MID	HI	LO	MID	HI
1985	5.93	6.64	7.76	5.61	6.32	7.44
1990	5.97	7.50	8.93	5.64	7.16	8.60
1995	6.08	8.18	10.87	5.73	7.82	10.51
2000	5.93	8.61	11.18	5.56	8.24	10.81
2010	6.04	8.49	10.32	5.67	8.13	10.95

Figure VIII.D.2. Comparison of Residential, Commercial, and Industrial Gas and Residential and Commercial Distiliate

are simply horizontal lines on the graphs. No adjustments have been made for transportation or distribution costs of the gasification plant products, exept for residental distillate. In this case $2/MMBTU was subtracted from the cost of residential distillate to reflect the cost differential between residential and well head gas (estimate obtained from EIA).

As can be seen from Figure VIII.D.3., MBG compares favorably with the mid-range price of competing fuels and should be highly competitive as an industrial boiler fuel.

While gasification plant methane can compete favorably only with the higher priced sources of new gas, Figure VIII.D.4. shows that it is highly competitive with distillate for space heating. Thus, as natural gas supplies decline, coal derived methane should be a competitive fuel in the space heating market.

Methanol and gasoline are compared in Figure VIII.D.5. Both products appear to be highly competitive with the mid-range forecasts. A significant point is illustrated by the figure; to the extent that methanol can be used as an above-average quality gasoline blending stock, it is always more economic to use methanol for blending than to convert it to gasoline. In other words, coal can be converted to gasoline more cheaply by blending methanol than by converting methanol to gasoline.

In summary, MBG, methanol, and gasoline appear to be highly competitive. Methane is only marginally competitive with the highest price competing fuels in the high-scenario forecast. Methanol is the most competitive alternate fuel, and is extremely attractive as a gasoline blending stock.

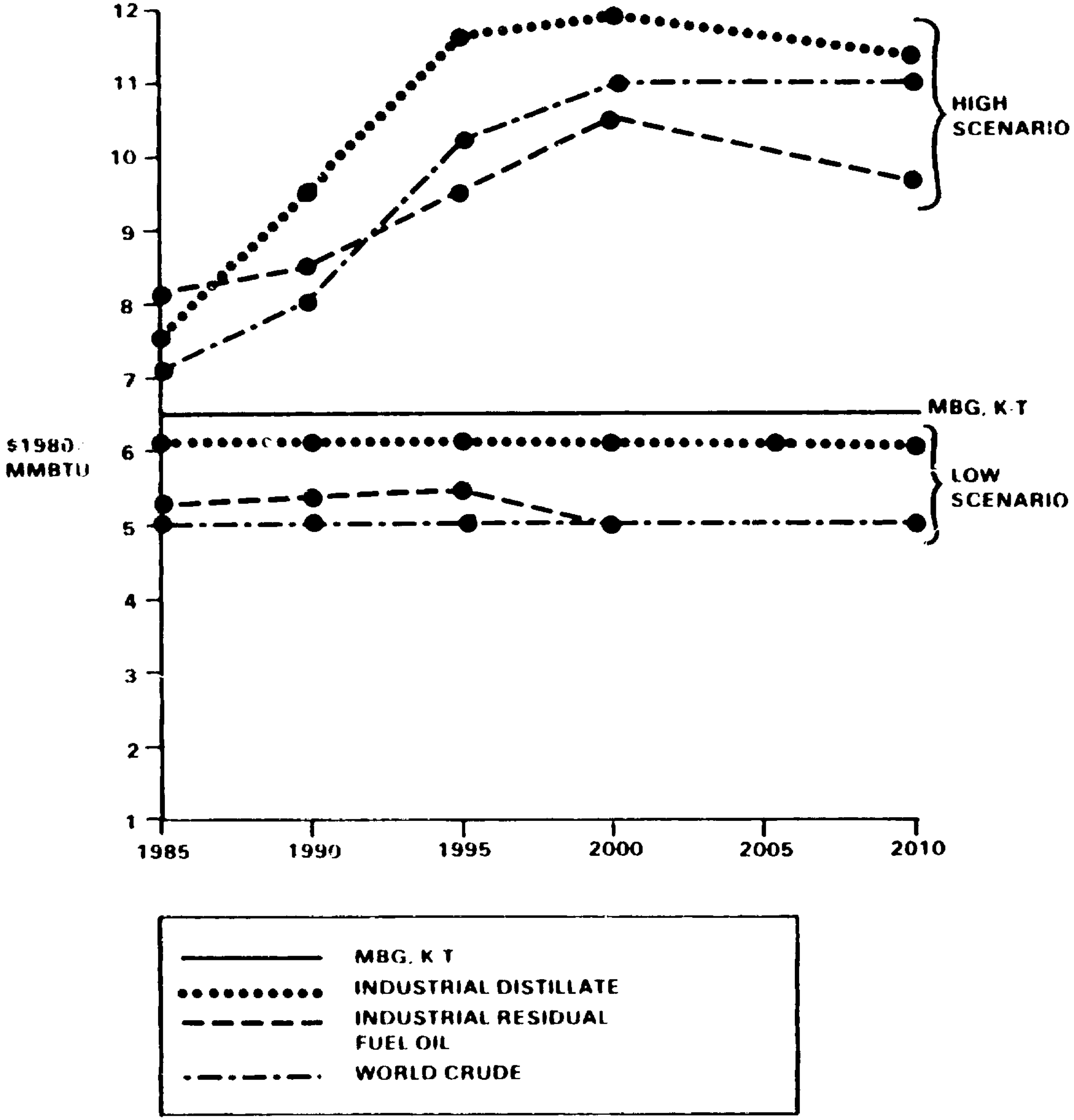
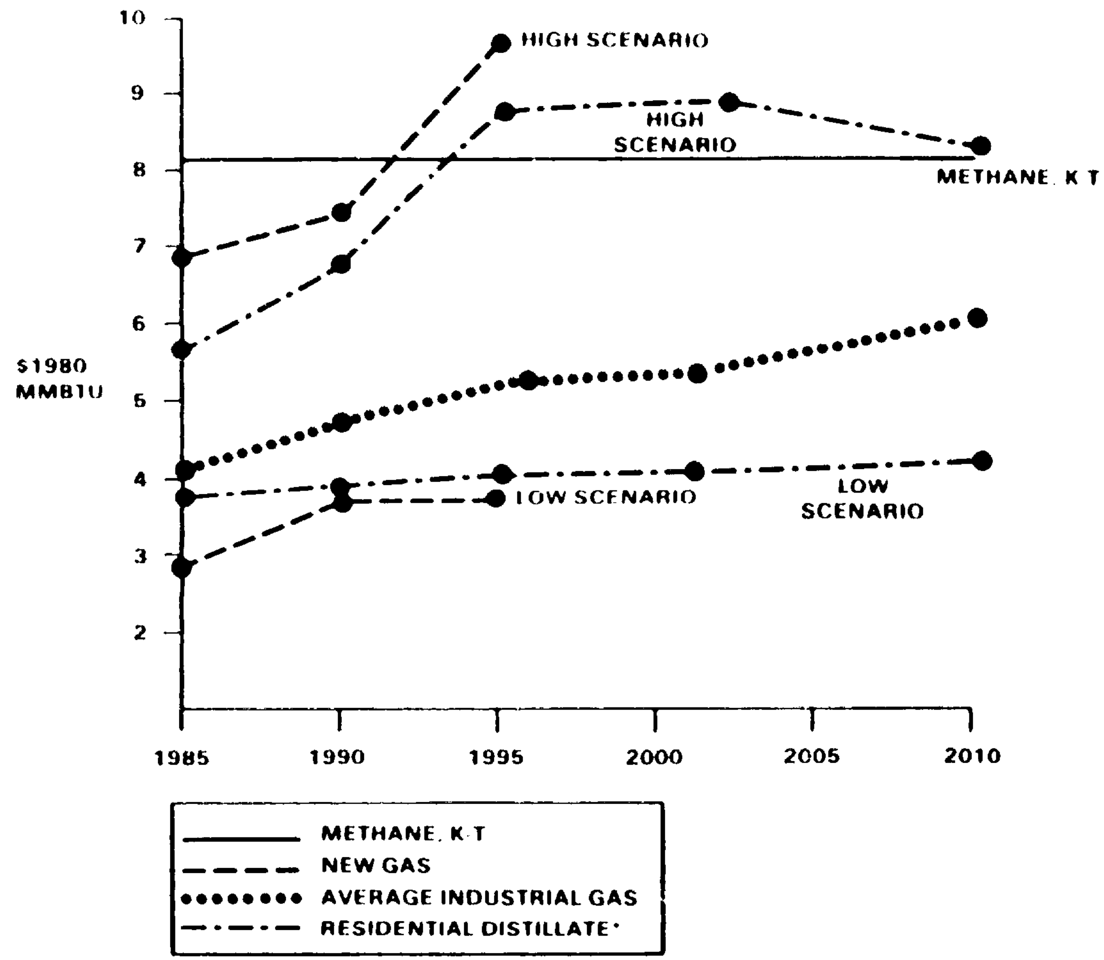


Figure VIII D.3. MBG Price Comparisons



*$2 MMBTU SUBTRACTED TO ADJUST FOR WELLHEAD DIFFERENTIAL WITH RESIDENTIAL GAS

Figure VIII.C.4. Methane Price Comparisons

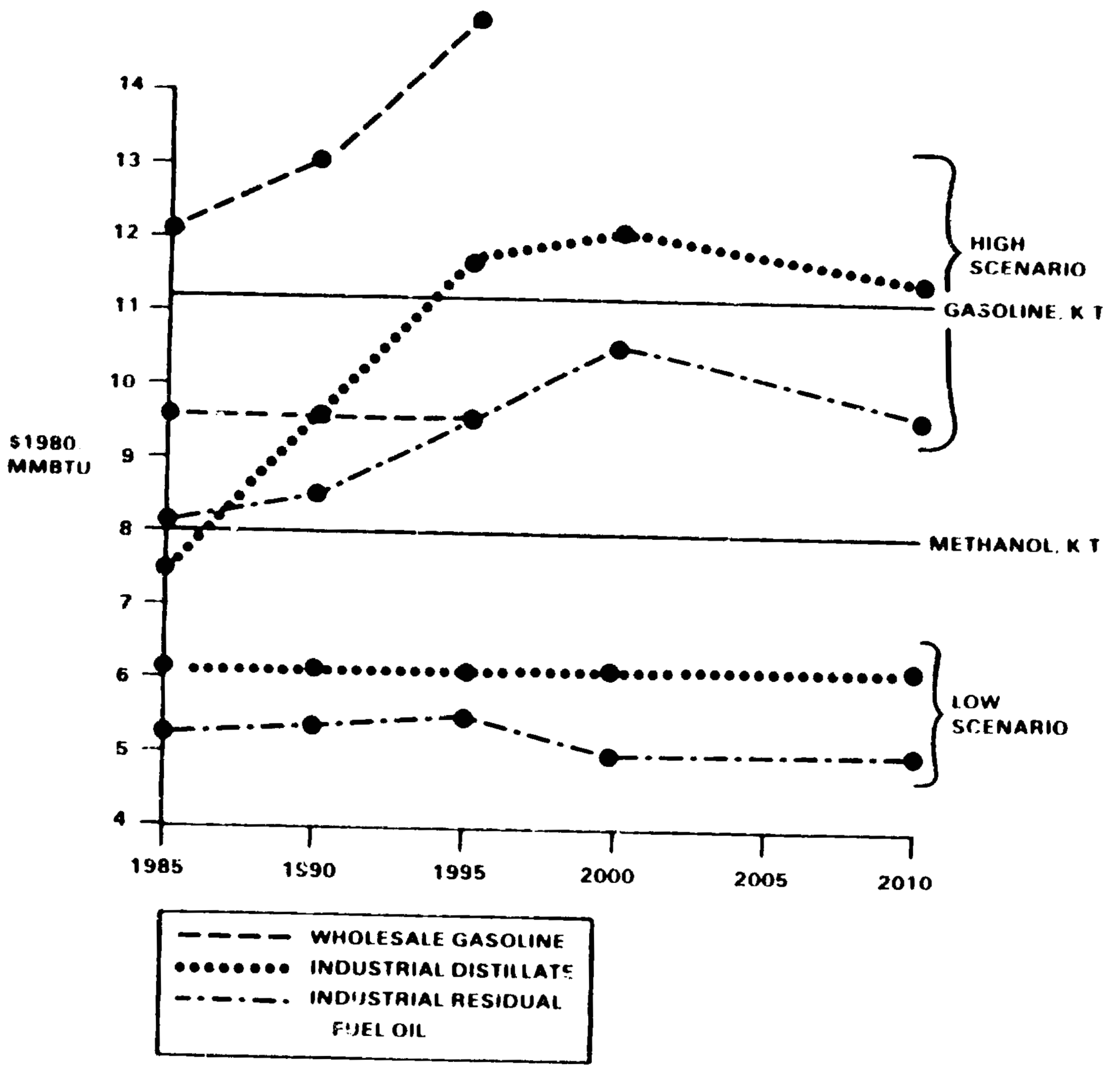


Figure VIII.D.5. Gasoline and Methanol Price Comparisons

	1980		1985		1990	
SCENARIO	$/BBL	$/MMBTU	$/BBL	$/MMBTU	$/BBL	$/MMBTU
LOW	33.43	5.57	29.64	4.94	29.60	4.93
MEDIUM	33.43	5.57	35.15	5.86	40.01	6.67
HIGH	33.43	5.57	42.27	7.05	48.30	8.05

	1995		2000		2010	
	$/BBL	$/MMBTU	$/BBL	$/MMBTU	$/BBL	$/MMBTU
LOW	29.61	4.94	29.59	4.93	29.59	4.93
MEDIUM	44.48	7.41	47.12	7.85	47.12	7.85
HIGH	60.97	10.16	65.76	10.96	65.76	10.96

Figure VIII.D.7 World Oil Prices, $1980/BBL

	INDUSTRIAL RESIDUAL FUEL OIL, $/MMBTU			INDUSTRIAL DISTILLATE, $/MMBTU			INDUSTRIAL POWER, $/KWH		
	LO	MID	HI	LO	MID	HI	LO	MID	HI
1980	3.88	4.61	5.42						
1985	5.34	6.31	7.46	6.16	6.95	8.18	4.16	4.27	4.27
1990	5.40	7.15	8.53	6.20	7.87	9.45	4.49	4.60	4.60
1995	5.50	7.84	9.55	6.31	8.60	11.56	4.38	4.49	4.49
2000	4.92	7.84	10.51	6.17	9.11	11.92	4.73	4.80	4.80
2010	5.06	7.77	9.69	6.29	8.98	10.98	4.59	4.64	4.64

	RETAIL GASOLINE, $/MMBTU			WHOLESALE GASOLINE, $/MMBTU (90% of retail)		
	LO	MID	HI	LO	MID	HI
1985	10.72	11.94	13.38	9.65	10.75	12.04
1990	10.76	13.00	14.43	9.66	11.70	12.99
1995	10.72	13.98	17.09	9.65	12.58	15.38
2000		--			--	
2010		--			--	

Figure VIII.D.8. Projected Petroleum Product Process

CHAPTER IX
COAL GASIFICATION FACILITY WORK BREAKDOWN STRUCTURE MATRIX AND DICTIONARY

A. INTRODUCTION

The WBS developed for this project is designed to allow a standard and logical format for estimating the Coal Gasification Facility project cost and schedule, while at the same time permitting cost and economic comparisons of Coal Gasification Facility Systems with alternate and competitive candidate systems for each segment of the process plant.

The complete WBS matrix and dictionary, included in their entirety in Appendix I, are summarized briefly in the following sections.

B. WBS MATRIX

The total WBS matrix shown in Figure IX.1 is a three-dimensional structure that shows the interrelationship of (1) the hardware elements dimension, (2) the phases and functions dimension, and (3) the elements of cost dimension. This latter dimension is not further developed at this time but is provided to show the overall expansion capability built into the WBS matrix. This dimension will become more important in later years when the Coal Gasification Facility project approaches a Phase III start and is defined to the extent that the elements of cost can be planned and estimated with realism.

There is, of course, the fourth dimension of time which cannot be graphically shown but must be considered also. Each entry on the other three dimensions varies with time, and it is necessary to know these cost values by year for budget planning and approval, and to establish cost streams for discounting purposes.

While a multiple-dimensional approach may at first appear unduly complex, it actually provides benefits that far outweigh any such concern. This structural interrelationship provides the capability to

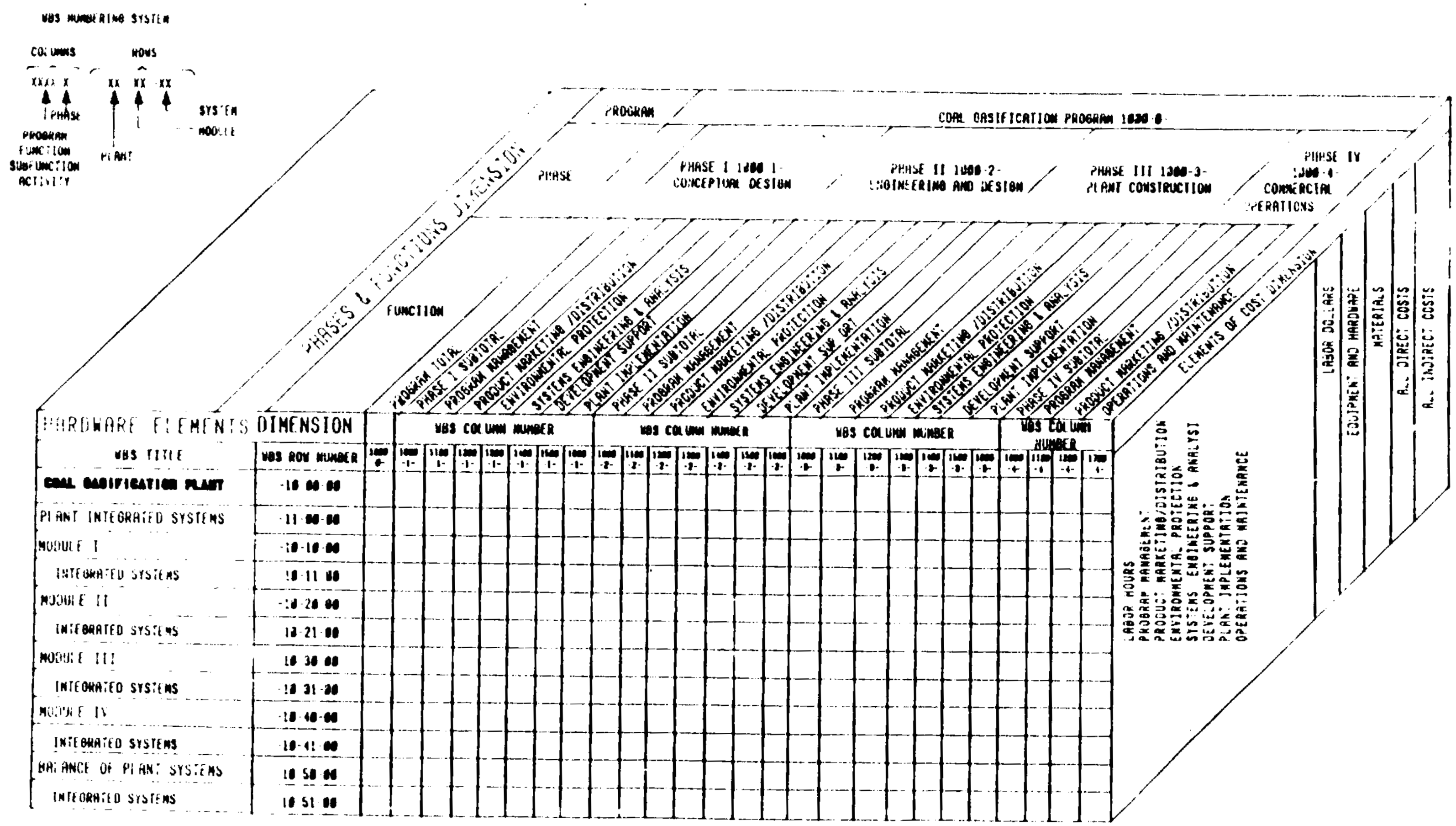


Figure IX.1. WBS Matrix

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view and analyze the Coal Gasification Facility from a number of different financial and management aspects. Costs may be summed by hardware groupings, phases, functions, etc. The WBS may be used in a number of three-dimensional, two-dimensional, or single-listing format applications.

C. DICTIONARY ORGANIZATION

The Coal Gasification Facility WBS Dictionary is divided into:

(1) A graphic display of three-dimensional WBS matrix

(2) The WBS hardware elements dimension of the matrix exploded into the standard graphic representation

(3) The WBS phases and functions dimension of the matrix exploded into the standard graphic representation

(4) The definitions of the hardware elements dimension

(5) The definitions of terms of the phases and functions dimension.

A systematic numerical coding system relates the rows of the hardware elements dimension to the columns of the phases and functions dimension, so that all matrix locations are identifiable by WBS numbers.

In Figure IX.1 a dot signifies each matrix position that corresponds to an identifiable task that must be completed in the Coal Gasification Facility program. Therefore, each dot also corresponds to a cost that will be incurred and must be accounted. Since each dot corresponds to one particular column of the accounts and phases dimension, a complete definition of any dotted matrix position is constructed by combining the definitions from the two applicable dimensions. That is, to avoid repetition, definitions are provided only once for each hardware elements dimension row and only once for each phases and functions dimension column, and a complete definition for any dotted matrix position is a combination of these two definitions.

D. HARDWARE ELEMENTS DIMENSION

The hardware elements dimension contains all of the presently defined hardware elements of the Coal Gasification Facility broken out

into facility, module, and system levels. Inherent within this dimension is the capability for further expansion to lower levels such as subsystem, assemblies, subassemblies, components, etc., limited only by the realism of the requirements.

E. PHASES AND FUNCTIONS DIMENSION

The phases and functions dimension is divided into five major phases: Each of these five phases is subsequently subdivided into functions such as program management, engineering, construction, test, development support, operations, etc., and each can be further subdivided into subfunctions such as design and development, systems engineering and analysis, manufacturing, site installation, system test and evaluation, and others.

CHAPTER X
MANAGEMENT POLICY AND PROCEDURES MANUAL

A. INTRODUCTION AND BACKGROUND

This report presents a review and recommendations for a TVA draft Management Policy and Procedures Manual[1] (MPPM) for the Coal Gasification Project (CGP). The draft was adapted by TVA from a similar manual for the Clinch River Breeder Reactor project.

To ensure successful completion, major projects require the early establishment of management policies and procedures manuals. In most organizations, it is customary to draw up a new manual for each project. The manual provides a clear understanding to all participants regarding their respective management and reporting responsibilities and procedures. The need for formal documentation is greatly enhanced to the extent the project has the following characteristics:

(1) Large
(2) Technology new to the project personnel
(3) New project organization and team of personnel
(4) Multiple participating organizations
(5) Uncertain definition of the end product of the project.

TVA's coal gasification project has all these characteristics. Additionally, it is a first-of-a-kind plant in the United States. Consequently, a MPPM is essential to assure effective integration of the organizational elements of the TVA CGP.

B. OVERVIEW OF MANAGEMENT POLICY AND PROCEDURES MANUAL

For most large projects, several levels of management plans are required. The Level I manual, outlined in Figure X.B.1, is the MPPM for the program manager. It defines the nature and organization of the overall

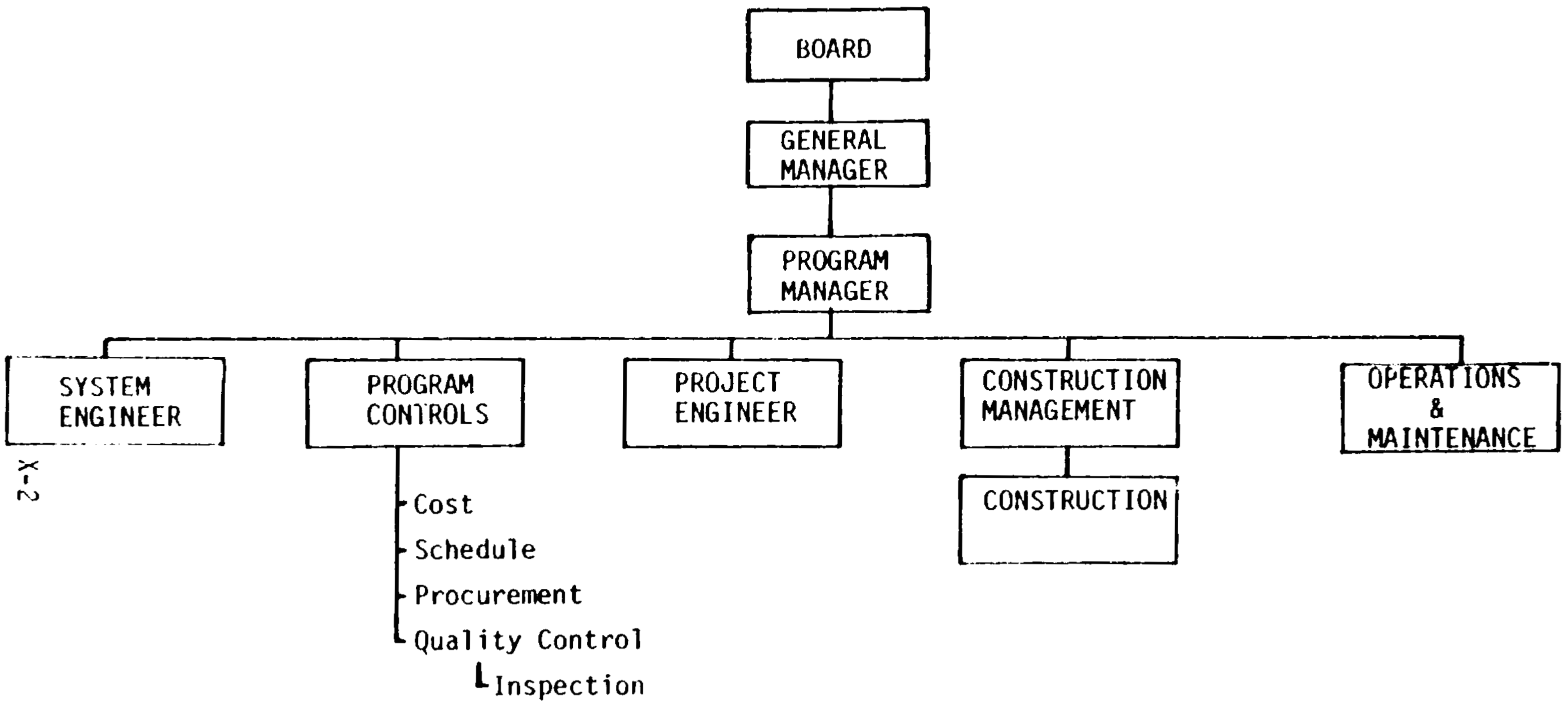

Figure X.A.1. Hypothetical Program Organization

project, describes the responsibilities of each organizational element, and lays out a master schedule and PERT chart for each major management activity. The Level II manuals establish the same information, plus detailed definition of management procedures, for each of the organization elements reporting to the program manager. Level III plans establish management and reporting procedures for the implementing organizations, particularly construction, operations and maintenance. The phase, function and subfunction categories are the same ones that appear in the Work Breakdown Structure described in Chapter IX. Figure X.B.2 illustrates a format that is applicable to all three levels for describing the assignment of responsibilities. Figure X.B.3 shows a detailed example of the bold outlined portion of Figure X.B.2 based on the hypothetical organization illustrated in Figure X.A.1. Figure X.B.4 provides a representative list of the management subfunctions that are typically included in the Level I manual. The Level I MPPM must be developed before the Level II and III manuals can be completed. Consequently, the Level I plan is the primary focus of this evaluation.

C. DESCRIPTION OF LEVEL I MPPM

The program definition section establishes the general program objectives, organization chart and staffing, background and description of the program, and program criteria or specifications. The description typically includes a brief general overview or summary of the major elements of the project such as the site, technologies, capacity and schedule. The criteria includes project specifications similar to those provided in the Design Criteria published by TVA in March, 1980. In the program activities section, each of the program activities listed in Figure X.B.4 is briefly defined and responsibilities are established as illustrated in Figures X.B.2 and X.B.3. Finally, a master schedule and PERT chart for the program is included. Figure 6 contains an example from BDM's systems engineering master plan for the Strategic Petroleum Reserve project.

1.0 PROGRAM DEFINITION

- 1.1 PROGRAM OBJECTIVES
- 1.2 PROGRAM ORGANIZATION
- 1.3 PROGRAM BACKGROUND AND DESCRIPTION
- 1.4 PROGRAM CRITERIA

2.0 MANAGEMENT TASKS

- 2.1 ENUMERATION AND DESCRIPTION
- 2.2 RESPONSIBILITIES

3.0 MASTER SCHEDULE & PERT OF MANAGEMENT TASKS

Figure X.B.1. Level I Management Plan Outline

WBS			ORGANIZATION ELEMENT/RESPONSIBILITY		
PHASE	FUNCTION	SUBFUNCTION	TVA GM	TVA PM	
I CONCEPTUAL	GENERAL MANAGEMENT	ADMINISTRATIVE	ME		
		PROGRAM DEVELOPMENT	S	ME Q	
	PROGRAM CONTROL				
	SERVICES				
II DETAIL DESIGN					
III FINAL DESIGN, CONSTRUCTION, START UP AND TEST					
IV OPERATIONS					

LEGEND

A Assist or Advise in Execution
E Direct Responsibility for Executing the Work
S Direct Responsibility for Supervising the Completion of the Work
Q Technical Responsibility for Quality of Work Produced
M General Management Responsibility for the Work
R Recipient of the Work Product (This Relationship Involves the Transfer of Information -- Not Any Coordinative Responsibility)

Figure X.B.2. Management Responsibilities

FUNCTION	SUB-FUNCTION	PROGRAM ORGANIZATION ELEMENT					
		SYSTEMS ENGINEERING	CONTROLS	PROJECT ENGINEERING	CONSTRUCTION MANAGEMENT	OPERA-TIONS	PROGRAM MANAGER
GENERAL MANAGEMENT	ADMINISTRATION PROGRAM DEVELOPMENT MANAGEMENT PLAN						
PROGRAM CONTROL	PRODUCT ASSURANCE QUALITY ASSURANCE MANAGEMENT CONTROL SYSTEMS DOCUMENT CONTROL & RECORDS MANAGEMENT						
SERVICES	PROCUREMENT ILS ENVIRONMENTAL INTEGRATED TEST TRAINING OPERATION & MAINTENANCE STANDARDIZATION R-A-M						

Figure X.B.3. Detailed Example of Bold Outlined Section in Figure X.B.2

PROGRAM ACTIVITY	EXAMPLES OF LEVEL I ACTIVITIES	EXAMPLES OF LEVEL II ACTIVITIES
GENERAL MANAGEMENT PROGRAM DEFINITION	PROGRAM OBJECTIVE CRITERIA DOCUMENTS PROJECT ORGANIZATION FACILITY CONCEPT	CONSTRUCTION CONTRACTORS DEFINITION CONSTRUCTION CONTRACTORS SELECTION CONTRACTS MONITORING AND INTEGRATION FACILITIES AND EQUIPMENT TESTING AND ACCEPTANCE INTEGRATED OPERATIONAL TEST AND ACCEPTANCE
ADMINISTRATION	ADMINISTRATIVE RESPONSIBILITIES PLAN	STAFFING PAYROLL ACCOUNTING PROJECT OFFICE FACILITIES
PROGRAM CONTROL MANAGEMENT CONTROL SYSTEM	POLICY AND GENERAL GUIDANCE FOR: -MILESTONE SCHEDULES -COST SCHEDULE -FINANCIAL CONTROL -CONFIGURATION CONTROL -PROCUREMENT CONTROL	PREPARE IMPLEMENTATION PLANS IN WITH GENERAL GUIDANCE PROVIDED BY PROGRAM OFFICE -DETAILED TIME AND COST SCHEDULE -FINANCIAL CONTROL -CONFIGURATION CONTROL -PROCUREMENT
DOCUMENTATION CONTROL AND RECORDS MANAGEMENT	DEVELOPS AND PROVIDES DOCUMENTATION CONTROL REQUIREMENTS	DEVELOPS AND IMPLEMENTS SPECIFIC REQUIREMENTS AND PROCEDURES FOR CONTRACTORS AND GOVERNMENT PERSONNEL TO FOLLOW
PRODUCT ASSURANCE	PRODUCT ASSURANCE PROGRAM PLAN	DEVELOPS IMPLEMENTATION PLAN TO INCLUDE: -CONTRACT PROVISIONS, DOCUMENTATION, INSPECTION SERVICES, PROGRAM INTERFACES AND CONTRACTOR REPORTS

Figure X.B.4. Illustrative Management Subfunctions

PROGRAM ACTIVITY	EXAMPLES OF LEVEL I ACTIVITIES	EXAMPLES OF LEVEL II ACTIVITIES
QUALITY ASSURANCE	QUALITY ASSURANCE PROGRAM PLAN -Q.A. CRITERIA	DEVELOPS IMPLEMENTATION PLAN TO INCLUDE: -SYSTEM/EQUIPMENT IDENTIFICATION PROCEDURES -CONTRACTOR REPORTING REQUIREMENTS -INSPECTION SERVICES -Q.A. CRITERIA
DEVELOPMENT SUPPORT SERVICES PROCUREMENT	ADVANCE BUDGET REQUIREMENTS BUDGET PLANNING SOURCE LIST	LONG LEAD PROCUREMENT DESIGN CONTRACT AWARD SITE PROCUREMENT PLANNING
INTEGRATED LOGISTICS SUPPORT	DEFINE LOGISTICS CONCEPTS FEASIBILITY STUDIES LIFE CYCLE/ECONOMIC COST	SYSTEMS LOGISTICS PLANS INTEGRATED LOGISTICS SUPPORT DEVELOPMENT SPARE PARTS PROVISIONING COST ANALYSIS
ENVIRONMENTAL	ISSUE NOTICE OF INTENT PREPARE EIA	PREPARE DRAFT SITE SPECIFIC EIS PREPARE SITE SPECIFIC EIS REQUEST PERMITS CONDUCT PUBLIC HEARINGS
STANDARDIZATION	IDENTIFY AREAS OF STANDARDIZATION INTERCHANGEABILITY SPECIFICATIONS	DEVELOP CRITERIA/PROCEDURES COMPILE SOURCE DATA
INTEGRATED TEST	ESTABLISH CRITERIA FOR MAJOR MAJOR SUBSYSTEM TESTING SYSTEM ACCEPTANCE TEST PLAN	DEVELOP TEST PROCEDURES PREPARE IMPLEMENTING PLANS DETERMINE RESOURCE REQUIREMENTS
TRAINING	PREPARE CRITERIA DOCUMENT AND OUTLINE GENERAL TRAINING REQUIREMENTS	DEVELOPMENT SPECIFIC TRAINING REQUIREMENTS BASED ON METHOD OF OPERATION AND O&M ANALYSES

Figure X.B.4 Illustrative Management Subfunctions (Continued)

PROGRAM ACTIVITY	EXAMPLES OF LEVEL I ACTIVITIES	EXAMPLES OF LEVEL II ACTIVITIES
OPERATION AND MAINTENANCE	OPERATION AND MAINTENANCE PLAN -PROVIDES GUIDANCE AND CRITERIA DEMONSTRATION PLAN	DEVELOPS IMPLEMENTATION PLAN TO INCLUDE: -OPERATION CONCEPT AND STAFFING AS REQUIRED -MAINTENANCE CONCEPT -PREVENTATIVE MAINTENANCE REQUIREMENTS -UNSCHEDULED/CORRECTIVE MAINTENANCE -O&M ENGINEERING ANALYSES PROCEDURAL ANALYSIS -O&M DOCUMENTATION -IMPLEMENTATION PLAN FOR DEMONSTRATION PROJECTS
MARKET PLANNING	PRODUCT EVALUATION PLAN	DETAILED MARKETING PLAN CONDUCT SPECIAL STUDIES
RELIABILITY AVAILABILITY MAINTAINABILITY	RAM PROGRAM PLAN -AVAILABILITY GOAL -RELIABILITY GOALS -MAINTAINABILITY GOALS	DEVELOP CRITICAL SYSTEMS RANKINGS AVAILABILITY ANALYSIS
ENGINEERING ENGINEERING	SELECTION OF ENGINEERING ORGANIZATION OR SUBCONTRACTORS	DETAILED ENGINEERING
CONSTRUCTION CONSTRUCTION	SELECTION OF CONSTRUCTION MANAGER AND CONSTRUCTION FIRMS SOURCE LIST	DETAILED CONSTRUCTION MANAGEMENT PLAN CONTRACTOR SELECTION AND MANAGEMENT

Figure X.B.4. Illustrative Management Subfunctions (Continued)

D. FINDINGS FROM REVIEW OF THE TVA DRAFT MPPM

The draft MPPM requires improvement in two respects. First, it has a considerable mixture of Level I, II and III policies and procedures within one document. The specific instances are documented in the annotated Table of Contents in Attachment A. Second, the draft does not completely treat any one of the levels. Most significantly, there is no assignment of responsibilities, and no master plan.

In large projects, it is important to develop separate level II and II manuals since each level represents major separate activities; and to do so with the intensive participation of the corresponding level II and III managers. Particularly in a new project organization, this will help to assure workable procedures based on the best collective thinking from the project organization. Based on these findings, the following recommendations were developed, rather than attempting to critique the draft MPPM in detail.

E. RECOMMENDATIONS

(1) TVA should develop a Level I Manual consistent with the description provided in sections B and C above, and consistent with the project Work Breakdown structure.

(2) The Level II managers should participate in development of the Level I manual. They should simultaneously develop outlines of the Level II manuals to ensure consistency between the Level I and Level II manuals.

(3) To ensure efficient project team operation, procedures consistent with existing TVA and contractor practices should be specified to the extent they are compatible with the project.

(4) After the Level I manual is completed, Level II and III manuals should be developed with the intensive participation of the corresponding Level II and III managers, and with guidance and approval of the program manager.

Attachment A

Annotated Table of Contents from TVA's draft MPPM, showing where Level I, II and III Management issues are addressed.

MANAGEMENT POLICIES AND REQUIREMENTS

TABLE OF CONTENTS

MANAGEMENT POLICIES AND REQUIREMENTS

TABLE OF CONTENTS (CONTINUED)

SECTION 2.5 - MANAGEMENT CONTROL SYSTEM

TABLE OF CONTENTS

SECTION 2.5 - MANAGEMENT CONTROL SYSTEM

TABLE OF CONTENTS

CHAPTER XI
COMMERCIAL DESIGN ASSESSMENT

Choices facing TVA or any other organization relative to implementing a coal gasification project requires any number of commercial level assessments. Included among others are assessments of available technology, alternate design approaches supplied by A/E firms and the potential for variation in raw material supply. This chapter presents the results of tasks performed in this contract relative to these items.

The first task is an assessment of comparative gasification technologies. Technologies explored here are limited to the five gasifiers which were the basis of design work in other tasks.

The second task is the development of comparative evaluation criteria for use in making selections from alternate design approaches. The criteria and formats developed here were done specifically for use with designs done for TVA by Bechtel, Foster Wheeler and C. F. Braun. However, they should be applicable to other projects with minor adjustments.

Third, a brief qualitative discussion of the impact of switching coal supply from the design coal, Kentucky No. 9, to a western sub-bituminous coal is presented.

Appendix G contains full reports on the results of the tasks.

A. COMPARISONS OF THE FIVE GASIFIERS SELECTED FOR EVALUATION BY TVA

1. Comparison of Design Characteristics

Table XI.A.1 presents the comparative design characteristics and parameters of the five gasification processes, based on one module for each process.

As shown in the table, the number of operating gasifiers required to gasify 5,000 tons per day of raw coal ranges from 2, for the B&W and the Lurgi/BGC gasifiers, to 8 for the K-T process. Thus, the capacity per gasifier ranges from 625 tons/day to 2,500 tons/day. Those capacities are not necessarily the maximum capabilities of the various gasifiers, but they probably approach the upper limit of their current capabilities.

TABLE XI.A.1. COMPARATIVE DESIGN CHARACTERISTICS OF TVA GASIFICATION MODULES
(BASIS: 1 MODULE FOR EACH PROCESS)

	B & W	K-T	TEXACO	LURGI	LURGI/BGC
OPERATING GASIFIERS	2	8	3	6	2
SPARE GASIFIERS	1	1	1	1	1
FEEDSTOCK COAL:					
Raw coal feed to module, T/D[a]	5,000	5,000	5,000	5,000	5,000
Dried coal feed to gasifiers:					
As is (dried coal), T/D	4,614	4,567	5,000[b]	5,000[b]	5,000[b]
As MAF coal, T/D	3,806	3,806	3,806	3,806	3,806
Raw coal/operating gasifier, T/D	2,500	625	1,667	833	2,500
OXYGEN (as 100 % O_2):					
T/D per module	3,779	4,033	4,188	2,061	2,031
lbs/lb of MAF coal to gasifers	0.99	1.06	1.10	0.54	0.53
GASIFICATION STEAM:					
T/D per module	392	678	401[c]	9,775	1,317
lbs/lb of MAF coal to gasifiers	0.10	0.18	0.12	2.57	0.35
lbs/lb of 100 % O_2	0.10	0.17	0.11	4.74	0.65
RAW GAS PRESSURE, psia	240	20	690	450	450
RAW GAS TEMPERATURE, °F	1,800	2,700[d]	2,500	950	610
COMBUSTION ZONE TEMPERATURE, °F	3,000	3,300	3,000[e]	1,900	3,000[e]

NOTES:

[a] Short tons (2,000 pounds) per day.

[b] Coal drying not required.

[c] Coal slurry water chemically converted during gasification.

[d] Before quenching for slag solidifcation. Entry to waste heat boiler, after quenching, is about 1,800 °F.

[e] Estimate, after allowance for heat loss and endothermic reactions.

It can be seen that the Lurgi and Lurgi/BGC gasifiers use considerably less oxygen than the other gasifiers. On the other hand, the Lurgi and the Lurgi/BGC gasifiers use much more steam than the others. It should also be noted that the slagging bottom Lurgi/BGC gasifier does not use as much steam as the non-slagging bottom Lurgi gasifier since the slagging gasifier does not require excess steam in order to maintain a low combustion zone temperature.

The Lurgi and Lurgi/BGC gasifiers have slowly descending, fixed beds. The B&W, the K-T and the Texaco gasifiers all have entrained beds and consequently operate at much higher raw gas exit temperatures.

The gasifier operating pressure levels range from 20 psia for the K-T process to 690 psia for the Texaco process.

In terms of their impact on plant costs and other factors, the relative effect on each of the design characteristics may be summarized as:

(1) Gasifier coal capacity: Higher coal capacities per gasifier reduce the number of gasifiers required, along with all of their related equipment and controls, and therefore reduce the overall plant costs

(2) Oxygen consumption: Higher oxygen consumptions require larger air separation units (to provide the oxygen) and therefore increase the overall plant costs

(3) Steam usages: Higher steam usage increases overall plant costs and results in more effluent waste water (contaminated process steam condensate) requiring more effluent water treatment and reuse or disposal

(4) Gasification pressure: Higher gasification pressure reduces the compression requirements of the end-product gas. However, higher gasifier pressures require higher pressure steam, more compression of the oxygen feed, and more costly coal feeding equipment. It is very difficult to generalize the overall cost impact of higher gasifier pressures, but it probably lowers plant costs

(5) Type of bed: The fixed-bed gasifiers (Lurgi and Lurgi/BGC) have a high inventory of coal in their beds which provides an inherent safety factor in the event of a coal feed failure while oxygen feed continues to enter the gasifier. The entrained bed gasifiers do not provide this inherent safety factor.

The fixed-bed gasifiers require the coal feed to be sized within a specific range so as to avoid gas channeling in the beds resulting from plugging of the spaces between coal particles by coal fines.

The entrained-bed gasifiers operate at much higher temperatures than the fixed-bed gasifiers and will, therefore, produce very little, if any, tars, oils, naphtha or phenols.

2. Comparison of Yield and Performance Characteristics

Table XI.A.2 presents the comparative yields and other performance characteristics of the five gasification processes, based on one module per process. The Lurgi and the Lurgi/BGC data in the table reflect a preliminary 'Facility Definition Design' only, whereas the three processes in the table reflect a detailed 'Facility Technical design'. Therefore, the Lurgi and the Lurgi/BGC data may not be completely comparative to the other three processes.

The yields of methane (CH_4) and the tar-oil-naphtha in the table reflect the previous observation herein that high-temperature, entrained bed gasifiers should produce little methane and essentially no tar-oil-naphtha.

The Lurgi and the Lurgi/BGC plants use considerably more steam than the other three plants and, therefore, produce more contaminated waste water. The ammonia recovery from the Lurgi and the Lurgi/BGC plants is a by-product of the need to treat and upgrade their waste waters for reuse implant as cooling water makeup.

The Lurgi and Lurgi/BGC gasifiers require a crushed and size-graded coal feed containing no coal fines. A part of the coal fines produced by crushing the raw coal is burned as boiler plant fuel and the remainder of the coal fines would have to be sold as a by-product. The tar-oil-naphtha by-products are also burned as boiler plant fuel.

TABLE XI.A.2. COMPARATIVE YIELD AND PERFORMANCE CHARACTERISTICS OF TVA GASIFICATION MODULES (BASIS: 1 MODULE FOR EACH PROCESS)

	B & W	K-T	TEXACO	LURGI	LURGI/BGC
PERCENTAGE OF COAL CARBON CONVERTED	97.46	95.00	98.98	99.02	99.52
PERCENTAGE OF COAL CARBON CONVERTED TO CH_4	0.00	0.59	0.73	13.55	10.86
T-O-N YIELD, wt % on MAF coal [a]	0.00	0.00	0.00	8.09	8.09
ENDPRODUCT MBG:					
Higher heating value, Btu/SCF	303	305	291	308	380
Gross MBG product, 10^6 SCF/D	275.0	249.4	271.1	289.5	239.0
Gross MBG product, 10^9 Btu/D	83.3	76.0	78.9	89.2	90.8
Net MBG product, 10^6 SCF/D	244.4	230.6	269.4	289.5	239.0
Net MBG product, 10^9 Btu/D	74.0	70.3	78.4	89.2	90.8
BYPRODUCT SULFUR, T/D [b]	185	183	184	177	179
BYPRODUCT AMMONIA, T/D	0	0	0	67	67
INPLANT FUEL USAGE:					
MBG, 10^6 SCF/D	30.6	18.8	1.7	0.0	0.0
T-O-N, T/D	0.0	0.0	0.0	308.0	308.0
Coal, T/D	0.0	0.0	0.0	996.0	313.0
Total fuel, 10^9 Btu/D	9.3	5.7	0.5	32.4 [c]	17.4 [c]
RAW WATER DEMAND, gpm:					
Boiler feedwater makeup	114	141	22	1400	105
Cooling water makeup	2848	1374	2056	585 [d]	205 [e]
Other users	-	-	77	115	105
Water treatment makeup (at 5 %)	148	76	108	105	20
Contingency (at 10 %)	310	159	227	220	45
Total raw water demand	3420	1750	2490	2425	480

NOTES:

[a] T-O-N is tar, oil and naphtha.

[b] Short tons (2,000 pounds) per day.

[c] Coal taken as 11,000 Btu/lb (HHV) and T-O-N taken as 17,000 Btu/lb (HHV).

[d] 1310 gpm of treated wastewater also used as cooling water makeup.

[e] 205 gpm of treated wastewater also used as cooling water makeup.

Of the three entrained bed gasifiers (B&W, K-T and Texaco), the K-T gasifier exhibits the lowest coal carbon conversion. The B&W design recovers and recycles most of the unburnt carbon (char) carried out of the gasifier with the raw gas. The Texaco process recovers most of the unburnt carbon (soot) carried out of the gasifier with the raw gas and recycles the recovered soot-water stream for reuse in the gasifier coal feed slurrying. The K-T design does not recover and recycle unburnt carbon carried out of the gasifier with the raw gas, which probably explains why the K-T process exhibits the lowest coal carbon conversion.

3. Comparison of Cost Factors

Table XI.A.3 presents the capital investment and the annual operating cost estimates for the five gasification processes.

As discussed above, the Lurgi and the Lurgi/BGC designs reflect a lower level-of-effort and may not be completely comparable to the other three cost estimates.

Of the three entrained bed gasification processes, the Texaco plant exhibits the lowest estimated capital cost as well as the lowest estimated operating cost.

The levelized life-cycle product prices and the detailed cost estimating methodology are presented and discussed in Appendix D.

B. COMPARATIVE EVALUATION CRITERIA

Risk management is a major element of these comparison criteria, especially in the areas of development schedule and plant operability. The evaluation and analysis of risk in most cases are very subjective and vary from client to client.

Each of the systems/facilities strengths and weaknesses should be evaluated in detail for subsequent selection of the most suitable systems/technologies. Estimates of the changes that would be required to make all systems acceptable should be documented.

TABLE XI.A.3. COMPARATIVE COST FACTORS OF TVA GASIFICATION PLANTS
(BASIS: 4 MODULES FOR EACH PROCESS)

	B & W	K-T	TEXACO	LURGI [a]	LURGI/BGC [a]
COAL GASIFIED, T/D	20,000	20,000	20,000	20,000	20,000
BOILER COAL, T/D	0	0	0	3,984	1,252
COAL FINES SOLD, T/D	0	0	0	6,456	9,188
TOTAL COAL FEED, T/D	20,000	20,000	20,000	30,440	30,440
TOTAL CAPITAL REQUIRED, MM $	3,347	2,371	2,091	2,747	2,061
O & M COSTS, MM $/YEAR	138	188	128	93	35
COAL, CATALYSTS & CHEMICALS, MM $/YEAR	181	181	181	274	274
TOTAL OPERATING COSTS, MM $/YEAR	319	369	309	367	309

NOTE: All cost factors are in 1980 dollars.

[a] These two cost estimates reflect a lower level-of-effort design and may not be completely comparative to the other three cost estimates.

1. Validation

Since design comparison is meaningless if the design is not correct, criteria have been established for validation of A/E conceptual designs. Prominent on the list of validation criteria are design data base, design and cost correctness, design feasibility, and compatibility of the design with the Integrated Facility Requirements. The design and cost drivers, identified in Appendix A, the cost data and methodologies in Appendix D, and the issues raised in the Critical Technology Assessments in Appendix F are essential in validating the designs.

a. Design Data Base

Each A/E conceptual design should be reviewed to determined whether the design base experimental data are acceptable.

b. Energy and Material Balance

The A/E conceptual designs should be checked to ensure each system mass-balance is within one pound per hour on both compounds and elements. Additionally, system energy balances should be checked for agreement to within 1%.

Where some systems or subsystems fail to satisfy these validation criteria, estimates of the changes that would be required to validate each should be used for the comparison step.

Upon confirmation of the energy and material balances, the A/E cost data should be validated using the costing and product pricing methodologies from Appendix D.

c. System Technical Feasibility

The technical feasibility of the A/E conceptual design should be evaluated as to whether the proper equipment has been selected and whether critical items have been spared. Each design should be reviewed to determine whether the system utilizes proven equipment in a configuration and/or service similar to those previously used successfully in the same or similar scale.

d. Compliance with Scope of Work and Integrated Facility Requirements Document

Each A/E conceptual design should be reviewed for compliance with the scope of work. A qualitative estimate of the impact of each deviation

from the Scope of Work and Integrated Facility Requirements Document quantities should be made.

2. Comparison

Comparison and evaluation follow the validation effort. The valid systems and facilities in the A/E designs should be compared, both with other A/E designs and with the Reference Facility Designs of Appendix B, for facilities employing the same gasification technology. Tables XI.B.1 lists the criteria used as a basis for comparison. Appendix G includes a tabular format for use in this task.

C. QUALITATIVE IMPACT OF GASIFYING SUB-BITUMINOUS COAL

The potential for interruption of coal supply is such that consideration needs to be given to the alternatives available in the event of such a disruption. One conceptual possibility is the importing of sub-bituminous coal from the Western United States. Such an alternative would have numerous impacts and perhaps be impossible without serious throughput consequences and major revisions to the plant's process train. The discussions below are qualitative in nature and intended only to point out some of the more significant impacts. The discussion is not exhaustive and a more definitive quantitative study would undoubtedly reveal other impacts not addressed here.

Additionally, these observations made concerning process impacts should not be interpreted to imply that the plant will actually run on sub-bituminous coal but at a throughput or efficiency penalty. This will depend upon the particular equipment put into the plant and the degree of flexibility designed into the plant. Rather, these discussions relate to areas of process constraint that the facility would face assuming that the plant functions on sub-bituminous coal feed.

1. Coal Preparation and Feeding

Grinding and pulverizing equipment requirements are likely to be different and require modification or additional investment. In the case of

TABLE XI.B.1. EVALUATION CRITERIA

I. SYSTEM PERFORMANCE AND RELIABILITY

A. MATURITY OF TECHNOLOGY
B. SCALE-UP REQUIREMENTS
C. COMPLEXITY
D. CRITICAL TECHNOLOGY ASSESSMENT
E. OPERATING REQUIREMENTS
F. FLEXIBILITY
G. REDUNDANCY

II. SYSTEM COST COMPARISON

A. TOTAL SYSTEM CAPITAL REQUIREMENTS
B. SYSTEM OPERATING COSTS

III. PLANT PERFORMANCE

A. GROSS COAL REQUIREMENT
B. NET COAL REQUIREMENT
C. NET MBG PRODUCED
D. IMPORTED ELECTRIC POWER
E. BY-PRODUCTS EXPORTED
F. CATALYST AND CHEMICAL CONSUMPTION
G. MISCELLANEOUS EXPORTS AND IMPORTS
H. FLEXIBILITY

IV. PLANT DESIGN RELIABILITY

A. MATURITY OF TECHNOLOGY
B. COMPLEXITY
C. REDUNDANCY OF HIGH-RISK COMPONENTS
D. CRITICAL TECHNOLOGY ASSESSMENT

V. PLANT COST

A. TOTAL CAPITAL REQUIREMENTS
B. NET ANNUAL OPERATING COST
C. UNIFORM ANNUAL EQUIVALENT PRODUCT COST

VI. ENVIRONMENTAL RELATED CRITERIA

A. MATURITY OF CONTROL TECHNOLOGY
B. EFFLUENTS POSE SITING LIMITS
C. BY-PRODUCTS POSE ENVIRONMENTAL HAZARDS

the Texaco process, size distribution is particularly important in maximizing slurry concentrations and the impact is likely to be especially significant. Likewise, western coal contains significant amounts of bound (non-free) water which will not contribute to the slurrying process but will absorb heat in the gasification reactor.

Much western coal has a moisture content exceeding that of the design coal. Thus, drying equipment would be inadequate.

2. Gasification

Switching to western sub-bituminous coal will have several impacts on the gasification section.

(1) Slagging gasifiers may have refractory problems with ash of differend composition and lower viscosity. Gasifiers such as B&W, which depend upon a solidified slag layer to protect the refractory would have to operate at lower temperatures to accommodate lower ash fusion temperatures

(2) Lower operating temperatures would mean slower reaction rates and either less throughput or less conversion. Higher reactivities would be offsetting, but the net effect would require quantitative assessment

(3) Lower operating temperature would generally mean less heat recovery and an inadequate steam system

(4) Fouling of WHB tubes by condensing alkali metal salts may occur.

3. Initial Gas Clean-up and Cooling

Quenching of gases with entrained slag particles may result in a buildup of alkali salts in the quench circulating loop and subsequent fouling.

4. Acid Gas Removal

The low sulfur content of western sub-bituminous coals will greatly reduce the sulfur removal requirement. However, requirements for control of the heating value of the product gas by carbon dioxide removal may limit any benefit from the low sulfur content particularly when the sulfur specification is 20 ppm.

5. Sulfur Recovery

Greatly reduced sulfur content will mean a troublesome CO_2/H_2S ratio. modification of the Claus plant probably will be required. Some sort of recycle system may be required to maintain a minimum hydrogen sulfide content.

6. Air Separation

The high oxygen content of sub-bituminous coals should lead to a decrease in gaseous oxygen requirements.

7. Utilities

The amount of process derives steam available may decrease with a switch to sub-bituminous coal.

CHAPTER XII
CRITICAL TECHNOLOGY ASSESSMENT

A. INTRODUCTION

Coal gasification is a relatively unknown technology to the U.S. What data are available on commercial scale systems are generally controlled by interest from outside the U. S. While small pilot units have demonstrated various process results for a wide variety of cases, such units cannot provide commercial reliability or performance data for large scale equipment.

Design data are currently primarily extrapolated from such industries as oil refining, chemical and steel making. Control and instrumentation systems are either conceptual or based on small units from which design scaleup data are often not the object of the program.

Technology development needs span all of the systems. Table XII.A.1 presents the technology development needs for the main processing units, Table XII.A.2 is a similar presentation for supporting units. While development needs are shown for a large number of systems, the most critical areas are associated with the gasifier itself and those systems which either feed the gasifier or directly receive products from the gasifier. Tables XII.A.2 through XII.A.6 define those areas of technology development needs.

Downstream processing units are closer in nature to commercial operations of the same or similar processes. However, the scale of coal gasification plants and the uncertainty associated with unfamiliar components which may be present in coal gasification process streams result in certain technical issues. Table XII.A.7 outlines these issues.

TABLE XII.A.1

TECHNOLOGY DEVELOPMENT NEEDS

MAIN PROCESS UNITS

	COAL FEEDING	GASIFICATION	GAS COOLING	GAS PURIFICATION	SHIFT CONVERSION*	METHANATION
PROCESS DESIGN						
DATA	X	X	X	X	X	
METHODS		X	X			
MECHANICAL DESIGN						
SCALE	X					
MATERIALS OF CONSTRUCTION	X	X	X		X	
EQUIPMENT RELIABILITY	X	X	X	X	X	X
PROCESS CONTROL						
DEVICES						
SENSING	X	X	X		X	X
CONTROL		X				
METHODS		X	X		X	X
OPERATIONS						
LONG (8000-HR) COMMERCIAL SCALE RUNS		X	X	X	X	X
BY-PRODUCT RECYCLE		X	X			

*DEVELOPMENT NEEDS LESS SEVERE THAN THOSE OF OTHER SYSTEMS

TABLE XII.A.2

TECHNOLOGY DEVELOPMENT NEEDS

SUPPORTING UNITS

	COMPRESSION	OXYGEN PLANTS	TAR/OIL SEPARATION	SULFUR RECOVERY	WASTEWATER TREATMENT	SOLIDS DISPOSAL
PROCESS DESIGN						
DATA			X	X	X	X
METHODS			X			
MECHANICAL DESIGN						
SCALE	X	X				
MATERIALS OF CONSTRUCTION	X		X			
EQUIPMENT RELIABILITY	X		X			X
PROCESS CONTROL						
DEVICES						
SENSING					X	X
CONTROL					X	
METHODS						
OPERATIONS						
LONG (8000-HR) COMMERCIAL SCALE RUNS						X
BY-PRODUCT RECYCLE			X			

TABLE XII.A.3

MATERIALS OF CONSTRUCTION

- CONVENTIONAL MATERIALS FOR PETROLEUM REFINING AND PETROCHEMICAL PLANTS MAY BE INADEQUATE IN COAL GASIFICATION PLANTS.
  - -- EROSION AND CORROSION MECHANISMS MAY BOTH BE IMPORTANT
  - -- MANY DIFFERENT CONSTITUENTS MAY BE PRESENT AT ONE TIME
- IN PARTICULAR, MANY ALLOYS EXHIBIT A TRANSITION TO RAPID CORROSION AFTER 1000 TO 5000 HOURS, AT HIGH H_2S CONCENTRATIONS CHARACTERISTIC OF COAL GASIFICATION ATMOSPHERES.
- AREAS OF HIGH-H_2O PARTIAL PRESSURE MAY SUBJECT CERTAIN REFRACTORIES TO DEGRADATION BY LEACHING OF SILICA FROM THE MATERIAL.
- MULTIPLE-LAYER MATERIALS MAY OFFER BETTER PROTECTION AGAINST COAL GASIFIER ENVIRONMENTS THAN ANY SINGLE SUBSTANCE.

TABLE XII.A.4

PROCESS CONTROL REQUIREMENTS

- DEPENDS ON
  - -- END USE
  - -- MULTIPLICITY OF TRAINS
  - -- TYPE OF GASIFIER
- RAPID SENSING OF FAILURES OR CHANGES IN PROCESS CONDITIONS
- POSITIVE CONTROL MEASURES
  - -- TURNDOWN/LOAD FOLLOWING
- DUPLICATION AND REDUNDANCY ARE ESSENTIAL
- COMPUTER GUIDED CONTROL MAY BE REQUIRED, ESPECIALLY FOR COMPLEX, CLOSELY COUPLED SYSTEMS

TABLE XII.A.5

SENSING AND MEASUREMENT REQUIREMENTS

- PRESSURES TO 1200 PSIG
- TEMPERATURES TO 3000°F
- FLOW OF GAS, LIQUIDS, AND SOLIDS
  - GASES CONTAINING H_2, H_2S, TARS, OILS, PARTICULATES CHLORIDES, ALKALIS AT TEMPERATURES TO 3000°F, PRESSURES TO 1200 PSIG
  - SOLIDS
    - COAL FEED TO GASIFIER
    - ASH OR SLAG FROM GASIFIER
    - TAR/OIL/FINES RECYCLE RATES
- PARTICULATE RATE AND SIZE CONSISTS
  - HOT DIRTY GAS
  - ADSORBED TARS AND OILS
- ON-LINE ANALYSIS
  - PRIMARY COMPOUNDS: H_2, CO, CO_2, H_2O, CH_4, N_2
  - POLLUTANTS: H_2S, COS, HCN, NH_3
  - CONDENSIBLES: PHENOLS, AROMATICS, OLEFINS, TARS, ACIDS
  - SOLID WASTES: METALS, ALKALI SALTS
- INVENTORY
  - BED LEVEL IN FIXED-BED AND FLUID-BED GASIFIERS
- SAFETY
  - IN-GASIFIER MONITORING OF OXYGEN BREAKTHROUGH
  - INTEGRITY OF REFRACTORY LININGS

TABLE XII.A.6

MEASUREMENT AND CONTROL REQUIREMENTS

- SENSING AND MEASUREMENT
  - -- PRESSURE IN GASIFIER RAW GAS STREAMS
  - -- TEMPERATURE INSIDE GASIFIERS
  - -- FLOW OF GASIFIER RAW GAS, FEED COAL, AND ASH
  - -- PARTICULATE CONTENT OF GASIFIER RAW GAS
  - -- ANALYSIS OF CONSTITUENTS OF GASIFIER RAW GAS
  - -- INVENTORY OF COAL WITHIN THE GASIFIER
  - -- GASIFIER SAFETY MONITORING

- CONTROL VALVES
  - -- PRESSURE LETDOWN ON SOLIDS AND SLURRY STREAMS
  - -- FLOW AND PRESSURE CONTROL ON HOT DIRTY GAS STREAMS
  - -- SHUTOFF VALVES FOR MULTI-TRAIN INSTALLATIONS

TABLE XII.A.7

TECHNICAL ISSUES IN DOWNSTREAM PROCESSING

- END-USES MAY DICTATE PRODUCT PURITY SPECIFICATIONS AND, THEREFORE UPGRADING REQUIREMENTS
  - -- COMBINED-CYCLE POWER PLANTS OR INDUSTRIAL BOILERS MAY HAVE LOAD-FOLLOWING REQUIREMENTS WHICH IMPOSE:
    - -- TURNDOWN REQUIREMENTS
    - -- PROCESS CONTROL REQUIREMENTS
  - -- FEDERA' , STATE, AND LOCAL REGULATIONS MAY DICTATE EFFLUENT TREATM NT REQUIREMENTS WHICH REQUIRE:
    - -- KNOWLEDGE OF DETAILED COAL COMPOSITION
    - -- KNOWLEDGE OF DISTRIBUTION OF COAL CONSTITUENTS WITHIN THE SYSTEM
- PROCESS DESIGN ISSUES FOR NON-GASIFICATION UNITS
  - -- IS THE PROCESS COMMERCIALIZED ON SAME OR SIMILAR FEEDS?
  - -- ARE DESIGN METHODS AVAILABLE?
  - -- ARE DESIGN DATA AVAILABLE?
- MECHANICAL DESIGN ISSUES
  - -- MATERIALS OF CONSTRUCTION
  - -- TEST PROGRAMS
  - -- PRIOR EXPERIENCE IN SIMILAR SERVICE

B. IDENTIFICATION OF CRITICAL TECHNOLOGY ITEMS AND ISSUES

The design experience of the study team combined with a survey of published design works, published reports on technology development, and personal communication with equipment manufacturers and research institutes has been used to compile a list of more than fifty-five critical technology items and issues. These items were identified based on their impact as defined in Table XII.B.1 Appendix F contains the Critical Technolgy report for this work.

C. HIGHEST PRIORITY ITEMS

In order to arrive at a set of recommended items for a development program, the list of identified items was reviewed relative to their relation to Table XII.B.1. For purposes of these evaluations:

(1) Minimum annual O & M savings correspond to 0.13 times the associated capital cost of achieving the savings.

(2) Cost effects of delta service factors were taken from the sensitivity analysis in Chapter V. A 90 percent service factor and a 1.0 percent improvement yield a 0.4 percent cost reduction.

(3) Improved efficiency takes the form of increased product from a fixed energy input.

The items identified as high priority items are discussed in Appendix F. The outstanding ones are disussed here.

(1) Gasifier refractory - Current practice is to either design for frequent repair or to operate with a solidified slag layer protecting the ash. Operating with the slag layer requires a high heat flux through the gasifier walls producing either low pressure steam in a jacket or increased capital costs associated with high pressure boiler tubes imbedded in the refractory.

TABLE XII.B.1

IMPACT OF CRITICAL TECHNOLOGY ISSUES:

1. DESIGN - Data is required to design the plant to meet specifications or improve plant design optimization.

2. COST REDUCTION -

   a. Initial Capital Cost - Technology development will reduce plant initial capital cost.

   b. Replacement Capital Cost - Technology development will reduce the cost per year of replacement capital items.

   c. Maintenance Costs - Technology development will reduce annual plant maintenance costs.

3. OPERABILITY -

   a. Product Specs - Technology development is required to ensure that the plant meets product specs.

   b. Emission Specs - Technology development is required to ensure that the plant meets emission specifications.

   c. On-Stream Time - Technology development will improve on-stream time.

   d. Efficiency - Technology development will improve plant energy efficiency.

   e. Safety - Technology development will improve plant safety.

(2) Waste heat boiler - Current practice is to use parallel redundant WHB's to accommodate frequent cleaning. Improved materials, especially ceramic-based materials, are needed. Heat transfer design data are needed.

(3) Materials of construction - Improved materials of construction are required to render valves in coal slurry or entrained flow service suitable for long operating periods. Current practice is to design with parallel systems isolatable for maintenance by a double block system.

(4) Slurry pumps - Improved materials and designs are required to develop a reliable long life slurry pump. Current practice uses high capital cost positive displacement pumps with provisions for frequent valve replacement.

(5) Slurry pumping - Pumps capable of handling more concentrated slurries will reduce the thermal penalty currently associated with water slurry feed systems. Injection and mixing of small amounts of coal with the slurry near the point of feed injection may be an acceptable approach.

(6) Instrumentation and control - Temperature monitoring and subsequent control are required for precise control of gasifier reactant flow. A control system combining measurement of feed flow and composition and reactor temperature will allow control of oxygen and steam flow in an optimum manner. Each molecule of O_2 in excess of that required converts two CO molecules to CO_2 with associated heat effects and loss of product. Likewise, too little O_2 results in unconverted carbon. Current practice is to monitor CO_2 levels and adjust reactant proportions accordingly.

D. RECOMMENDED PROGRAM

Generally, the benefit from major development efforts in the process industries is derived by application of results over a broad operation of the applicable industry. Developments in coal gasification are likely to fall into this pattern also. Investments in coal gasification technology development programs are not normally project specific and are justified more easily by considering the total national potential than by considering the TVA project alone. However, the emphasis in this project has been on the TVA project and thus little consideration has been given to critical technologies specific to the Lurgi or BGC/Lurgi type gasification systems.

The most significant critical technology items are found to relate to the gasifier itself, the gasifier reactant feed system, and the recovery of heat from product gases. Benefits from these potential improvements take the form of improved service factors or improved efficiency. Up to 75 million dollars in development and capital costs are justified in improving efficiency by one percent in a single 20,000 TPD plant; up to 18 million dollars are justified in improving the service factor by one percent.

It is recommended that any coal gasification technology development program at MSFC have a large commitment to improving gasifier refractory improvement. Improvements in this area could benefit both service factors and efficiency. Excessive downtime to replace or repair refractory is costly. Avoiding refractory problems by operating with a solidified slag coating in the reactor requires either capital investment to imbed steam coils in the refractory or production of low pressure steam of marginal value in reactor jackets. A program to improve refractory is believed to have the greatest potential for major direct application in the TVA plant. The potential for retrofitting enhances this conclusion.

In order to establish additional potential for a major improvement in coal gasification technology, it is recommended that a large test facility be established suitable for developmental and test work on

prototype heat recovery and gas cleanup equipment. The capacity of this facility should be equivalent to several hundred tons per day coal feed in order to demonstrate flow similarities with large prototype or full scale equipment. The recommended approach to supplying this type of facility is to establish a slipstream or dedicated gasifier in conjuction with the TVA plant. If this proves not to be feasible, a test facility based on oil gasification should be established at MSFC. Oil gasification with injection of ash and other appropriate substances is preferred over coal in order to facilitate long term (months) testing and eliminate coal handling as a concern. Recycle of gas product would be used to minimize oil consumption and product disposal problems and at the same time provide a test facility for such things as gas compression prototype seal testing. Such a facility would also have potential for gasifying dilute but solids containing solvent refined coal as a means of duplicating coal ash concentrations.

Developmental efforts in rotating equipment and control valves have potential for savings in service factors, capital investment and maintenance costs. It is recommended that NASA efforts in this area be closely coordinated with equipment manufacturers and other existing test facilities such as the hot gas test loop at DOE's Morgantown Energy Research Center.

Better control of gasifier operation with improved efficiency and safety is especially valuable in a variable load situation which may develop for the TVA plant. A program to investigate plant dynamics and control strategy is recommended.

Finally, Table XII.D.1 is a summary of other technical items/issues, their impacts and an estimate of the resources required to address them.

The developmental program recommendations presented here are directed at concentrated efforts in relatively firm areas of hardware development and instrumentation and control. Process design data needs are significant but are not part of the recommended program other than the collection and analysis of data which my fall out of other programs.

TABLE XII.D.1 CRITICAL TECHNOLOGY ITEMS/ISSUES

SUBSYSTEM	CRITICAL ITEM	TECHNICAL STATUS	KEY ISSUE	ACTION	TIME REQ. (MONTHS)	COST, $ x 10^5
COAL PREPARATION	DRYING	CURRENTLY USES HEAT FROM COMBUSTION OF COAL	ENERGY CONS. & ECON.	BETTER DRYING TECHNIQUES (MICROWAVES)	6 24	LAB SCALE - 0.2 PILOT SCALE - 5.0
GASIFICATION	AGGLOMERATION WITH CAKING COAL (LURGI)	TESTING IN PROGRESS	OPERABILITY	TECH ASSESSMENT	6	0.2
	CONTINUOUS DRY COAL FEEDING	UNDER DEVELOPMENT	OPERABILITY & ECON.	TECH ASSESSMENT	4	0.1
				DEVELOPMENT PROGRAM	36	4.0
	PILOT PLANT TESTING ON PLANT COAL	NOT YET DONE	OPERABILITY & SCHEDULE	TEST PLANT COAL	10	3.0
	GASIFICATION CATALYSIS	SOME DEVELOPMENTAL WORK	ECONOMIC	TECH ASSESSMENT	4	0.15
				DEVELOPMENTAL PROGRAM	40	8.0
	MATERIALS RELIABILITY	SOME DEVELOPMENTAL WORK	OPERABILITY	TECH ASSESSMENT	6	0.2
				PILOT PLANT PROGRAM	20	4.0
GAS QUENCH, WASTE HEAT RECOVERY, AND COOLING	DISTRIBUTION OF GASEOUS AND TRACE METAL COMPS.	LITTLE DEVELOPWORK	SCHEDULE & OPERABILITY	METHODS EVALUATION	6	0.3
				PILOT SCALE TESTING USING NASA ROCKET TEST STANDS	36	5.0
	MATERIALS RELIABILITY	SOME DEVELOPMENTAL WORK	OPERABILITY	TECH ASSESSMENT	6	0.2
				PILOT PLANT PROGRAM	20	4.0
				MODULE #1 TEST PROGRAM	72	3.0
ACID GAS REMOVAL	COS REMOVAL	SOME TEST WORK	OPERABILITY & SCHEDULE	TECH ASSESSMENT	6	0.2
				HYDROLYSIS TEST PROGRAM	12	4.0
	SOLVENT DEGRADATION	LITTLE TEST WORK	OPERABILITY & ECON.	LABORATORY TESTING	12	0.4
				MODULE #1 TEST PROGRAM	60	2.0
	DISTRIBUTION OF ORGANICS	LITTLE TEST WORK	OPERABILITY	LABORATORY TESTING	12	0.4
				MODULE #1 TEST PROGRAM	60	2.0
ALTERNATE PRODUCTS	CATALYST POISONING	SOME TESTING	OPERABILITY & ECON.	LABORATORY TESTS	12	1.5
				MODULE #1 TESTS	36	6.0
	PRODUCT UTILIZATION (METHANOL IN GASOLINE)	SOME TESTING	ECONOMICS	TECH ASSESSMENT	6	0.2
				PILOT TESTS	24	3.0
SULFUR RECOVERY	LEVEL OF MINOR COMPONENTS HCN, COS, CS_2, RSH, H_xC_x	OPERATING PLANTS WITHOUT COMPONENTS	OPERABILITY & SCHEDULE	DEFINE QUANTITY	6	1.0
				MODULE #1 TESTING	12	6.0
	MATERIALS RELIABILITY IN SO_2 REMOVAL	SEVERAL TEST PROGRAMS	OPERABILITY	TECH ASSESSMENT	6	0.2
				MATERIALS TESTING	24	4.0
WATER TREATING	NITROGEN REMOVAL	SOME TESTING	SCHEDULE & OPERABILITY	LABORATORY TESTS	12	0.5
				MODULE #1	12	2.0
	TRACE ORGANIC REMOVAL	LITTLE TESTING	SCHEDULE & OPERABILITY	LABORATORY TESTS	12	0.5
				MODULE #1 TESTING	12	1.0
	BUILDUP OF MINOR COMPONENTS (CHLORIDE)	LITTLE TESTING	OPERABILITY	WATER SYSTEM MODELING	9	0.6
				MODULE #1 TESTING	24	1.0
	TRACE METALS IN WASTES	SOME TESTING	SCHEDULE	TECH ASSESSMENT	6	0.3
				PILOT PLANT TESTS	12	[illegible]
				MODULE #1 TESTS	24	[illegible]
SOLID WASTE [illegible]	LEACHABILITY OF METALS	SOME TESTING	SCHEDULE	LABORATORY LEACHING TESTS	[illegible]	[illegible]
				MODULE #1 TESTS	[illegible]	[illegible]

E. RESOURCE REQUIREMENT

The resource requirements for program discussed here are substantial. A hot test bed capable of melting coal ash and exposing refractory samples is required. The unit should be capable of operating under controlled large H_2O vapor pressures and oxidation/reduction environments. The design should accommodate testing of temperature measurement devices.

As stated above the recommended approach to establishing a large test bed for prototype testing of new hot gas process equipment is to establish it as a test facility at the TVA plant. The facility for an alternate approach based on oil gasification is described in Appendix F.

Manpower requirements for conducting a refractory development program can vary extensively depending upon the level of effort desired. As a minimum in order to conduct a meaningful effort two scientist/engineers knowledgeable in refractory and temperature measurement technology plus two lab technicians should be committed to the program.

Depending upon the final size selected, it is anticipated that the installation of a major test facility such as this will cost on the order of 20 to 50 million dollars. A staff of 30 to 40 persons would be required to support such a facility. If such a facility is built, it is recommended that a commercial supplier such as Texaco or Shell be contracted to furnish the design for the basic gasifier system.